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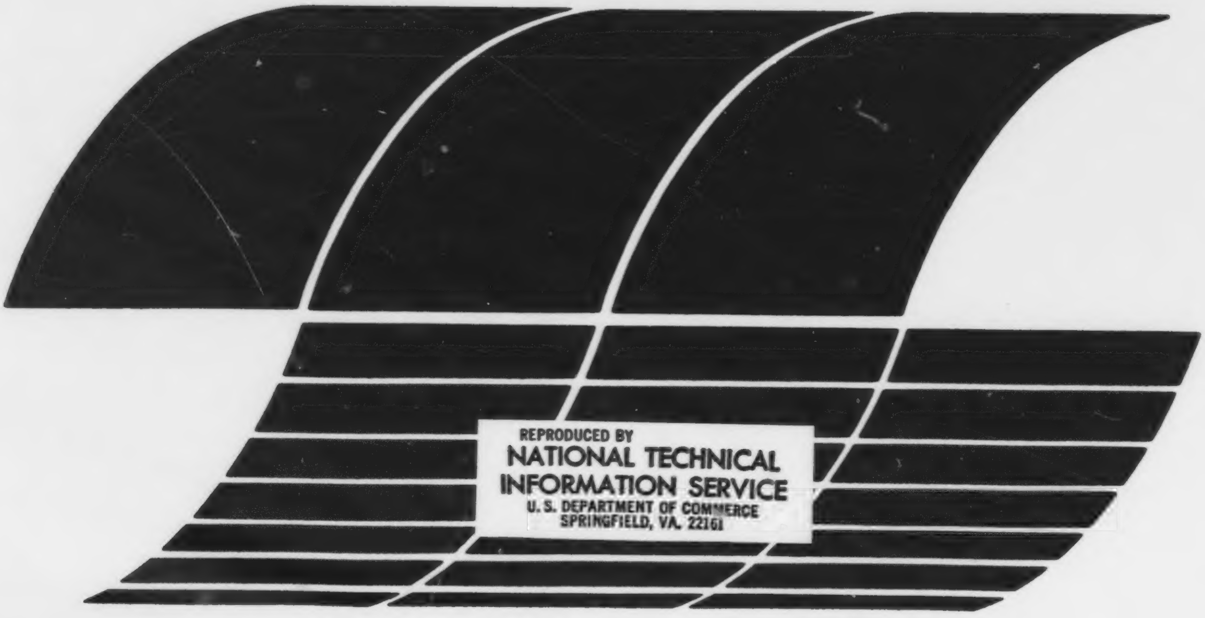

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16. ABSTRACT

The proceedings document presentations made during the August 29, 1978 industry briefing conference which dealt with the status of EPA/IERL-RTP's flue gas desulfurization (FGD) research, development, and application programs. Subjects considered included: lime/limestone scrubbing test results, forced oxidation, process cost and energy requirements, by-product disposal options, and future test plans. The conference provided developers, vendors, users, and those concerned with regulatory guidelines with a current review of progress made in IERL-RTP's FGD technology development program.

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Proceedings: Industry Briefing on EPA Lime/Limestone Wet Scrubbing Test Programs
(August 1978)

John E. Williams, Conference Chairman

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Industrial Environmental Research Laboratory
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PREFACE

More than half of all "man-made" sulfur dioxide (SO_2) is emitted by electric power plants, and the use of sulfur-containing fossil fuels, especially coal, to generate electricity is expected to increase dramatically in the next 10 years. Therefore, the develoment and commercial application of SO_2 control technologies is one of the most important concerns of the U.S. Environmental Protection Agency (EPA). Flue gas desulfurization (FGD) is the most promising technique for control of SO_2 that will be available for widespread application to fossil fuel-fired electric electric power plants for at least the next 8 to 10 years.

The Industrial Environmental Research Laboratory - Research Triangle Park (IERL-RTP) of EPA's Office of Research and Development periodically sponsors symposia and Industry Briefing Conferences for the transfer of information regarding FGD research, development and application activities with the objective of further accelerating the development and commercialization of this technology. One of the major IERL-RTP FGD efforts for the past several years has been advancement of the technology for lime/limestone wet scrubbing. The focal point of this program has been the prototype testing at EPA's Alkali Wet Scrubbing Test Facility, located at TVA's Shawnee Steam Plant, near Paducah, Kentucky. Current emphasis of the test program at Shawnee is to optimize lime and limestone systems in the areas of improved sludge disposal, performance reliability, and process economics.

The August 1978 Industry Briefing Conference focused primarily on recent test results at Shawnee in which the predominantly calcium sulfite reaction products were forced oxidized to calcium sulfate (gypsum). Potential advantages of forced oxidation were also discussed. Other material presented during the Conference included: IERL-RTP in-house pilot plant results, which have contributed significantly to an understanding of the lime/limestone system chemistry and in supporting the Shawnee program; a review of the IERL-RTP effort in the waste solids disposal area, including discussion of waste disposal economic options; and a review of future plans, including testing at TVA's Widow's Creek Plant and the EPRI test programs.

More than 125 people representing electric utilities, process suppliers, and State and Federal regulatory agencies, attended the Conference. The Chairman of the August 1978 Industry Briefing Conference was John E. Williams, a Chemical Engineer in the Emissions/Effluent Technology Branch, IERL-RTP.

These Proceedings are comprised of copies of the participating authors' papers as received. As supplies permit, copies of the Proceedings are available free of charge and may be obtained by contacting IERL-RTP's Technical Information Coordinator, Environmental Protection Agency, Research Triangle Park, North Carolina 27711.

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SIGNIFICANT EPA/IERL-RTP PILOT PLANT RESULTS

by

Robert H. Borgwardt

For Presentation at EPA Industry Briefing Conference

August 29, 1978

INTRODUCTION

The last review of IERL-RTP pilot plant testing was made at the FGD Symposium in November, 1977. It discussed the effect of forced oxidation on the performance of a single-loop limestone scrubber, showing that complete oxidation of the slurry could be accomplished at normal pH (to at least as high as 6.5) without adversely affecting SO_2 removal efficiency or scaling potential. Because of its distinct advantages with respect to sludge quality and operating simplicity, this scrubbing configuration has continued to be the focus of testing at RTP. Recent work with the single-loop limestone scrubber has involved two main areas of investigation: 1) the use of adipic acid as an additive for improving SO_2 removal efficiencies while forcing oxidation, and 2) the replacement of makeup water with simulated cooling tower blowdown as a means of further reducing the fresh water requirements for FGD and improving the overall water management in a power plant. This presentation reviews progress toward those objectives.

Adipic Acid

This prospective scrubber additive is a solid, straight-chain dicarboxylic acid: $HOOC(CH_2)_4COOH$. It was obtained in 50 lb bags at a cost of 46¢/lb for the tests at RTP. Bulk shipments are quoted at 41¢/lb.

The tests were undertaken as a result of theoretical analyses carried out by G. Rochelle ("The Effect of Additives on Mass Transfer in $CaCO_3$ and CaO Slurry Scrubbing of SO_2 from Waste Gases," Ind. Eng. Chem., Fundam. 16, pp. 67-75, 1977) which indicated that adipic acid should act as a buffer to limit the drop in pH that normally occurs at the gas/liquid interface during SO_2 absorption. The additional capacity of the surface film for SO_2 absorption brought about by this buffering action is expected to enhance the liquid-phase mass transfer and improve the overall SO_2 removal efficiency of a limestone or lime scrubber of a given type operating at a given L/G. The relatively low cost of adipic acid, together with its particularly favorable ionization constant, make it a prime candidate for testing as a buffering additive.

Further analysis by Rochelle ("Process Alternatives for Stack Gas Desulfurization by Throwaway Scrubbing," Proceedings of 2nd Pacific Chem. Eng. Congress, Vol. I, p. 264, August 1977) showed that additives will be most effective, cost-wise, when used in scrubbers employing forced oxidation. This approach would minimize the loss of the additive--and thus reduce the amount required for makeup--because of the tighter loop resulting from the better dewatering properties of oxidized sludge. For similar reasons, additive losses will also be minimized when fly ash is collected dry rather than collected in the scrubber. A fly ash-free scrubber producing oxidized sludge should achieve the greatest SO_2 removal efficiency for a given amount of additive.

Adipic acid has several potential advantages over other additives, such as MgO, that function by increasing the dissolved alkalinity (primarily as soluble sulfites). First, since adipic acid functions by a different reaction mechanism which does not include the sulfite/bisulfite equilibrium it is not affected by oxidation of the sulfite in the scrubbing liquor. It is therefore particularly well suited for use in a single-loop scrubber employing forced oxidation in the hold tank. Second, the buffering mechanism by which adipic acid enhances SO_2 absorption is not affected by the presence of chloride: the effectiveness of the alkali additives is reduced in proportion to the concentration of chloride. The lack of interference by chloride thus favors the use of adipic acid in systems employing forced oxidation since these systems concentrate chloride to much higher levels in the scrubbing loop. This feature also has importance with regard to the possible application of filter washing to remove the soluble salts from sludge--washing will also tend to concentrate chloride in the scrubber.

A third important advantage of adipic acid is indicated by initial cost comparisons made by Bechtel. Assuming that no additive losses occur in the solids, the effectiveness of adipic acid is expected to be sufficiently greater than that of MgO to more than compensate for its higher cost per pound (bulk MgO = 11¢/lb): the cost of obtaining a given degree of improvement in SO_2 removal should be lower for adipic acid.

The purpose of the IERL-RTP tests was to verify the postulated effect of adipic acid on SO_2 removal efficiency in a single loop limestone scrubber and to determine whether any deleterious effects are associated with its use, particularly with regard to sludge properties and oxidation efficiency.

Water Reuse

The other area of testing at IERL-RTP involved the substitution of a simulated cooling tower blowdown for the makeup water. The successful application of forced oxidation within the scrubbing loop is expected to facilitate the improvement of overall water reuse in a power plant by making possible the use of waste water from the boiler and/or cooling tower as FGD scrubber feed. The basis for this assumption is that the high density of pure gypsum in the oxidized slurry can rapidly dissipate the supersaturation occurring when the extra sulfate present in the blowdown (as Na_2SO_4) is fed to the scrubber.

The tests are ultimately aimed at the development of a scheme whereby a water treatment process, such as vapor compression evaporation, is incorporated within the FGD scrubber system. The soluble salts, including Na_2SO_4 and calcium chloride, would thus be simultaneously extracted from the scrubbing loop while regenerating enough fresh water for mist eliminator washing. This approach has the potential of eliminating soluble salts from the waste sludge and improving the energy efficiency of water treatment, while reducing the overall fresh water requirements. When combined with the use of adipic acid, such a system would have the further advantage of maximizing, the SO_2 removal that can be achieved with a given amount of additive. By also incorporating filter cake washing and additive recycle, the elimination of additive makeup is conceivable since the relatively insoluble adipic acid could be separated from the soluble salts during evaporation. The tests reported here are a first step in evaluating the feasibility of this concept.

RESULTS

Using the scrubbing configuration indicated by Figure 1, tests were made with two absorber types, a TCA and a multigrid (simulating a spray tower), operating at L/G = 65 gal/10^3cf (8.7 liters/m^3). Adipic acid was fed with the limestone. Figure 2 summarizes the effect of adipic acid on the SO_2 removal efficiency in the IERL-RTP pilot plant. At an adipic acid level of about 1600 ppm, the SO_2 removal efficiency of the TCA was increased from 82 to 93 percent. As indicated by Table 1, the limestone utilization was also significantly better than that obtained without adipic acid when compared at similar feed stoichiometry.

As expected, the SO_2 removal efficiency obtained with high chloride concentrations in the scrubbing liquor was not significantly different from that obtained without chloride when operating at similar levels of adipic acid.

The improvement in SO_2 removal could be maintained at the 85 percent limestone utilization level required for minimum mist eliminator fouling.

An oxidation rate of 1.3 x 10^{-3} g mol/1 (min) was maintained in the system at pH 5.9 using an unstirred, air-sparged tower of 18 ft (5.5 m) slurry depth. No significant reduction in oxygen transfer efficiency was observed at adipic acid levels up to 5000 ppm.

TABLE 1. Effect of Adipic Acid on TCA Scrubber Performance at Constant Limestone Feed Stoichiometry

	No Additive	Adipic Acid Added
SO_2 Removal, %	82	93
Limestone Utilization, mol %	77	91
Chloride Concentration, ppm	20,000	17,000
Scrubber feed pH	6.1	5.9
Scrubber effluent pH	5.1	4.9
Oxidation[a], mol %	99	99
Settling rate, cm/min	3.4	2.7
Limestone stoichiometry[b]	1.07	1.01
Adipic acid conc., ppm	0	1570

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[a] Air stoichiometry = 2.7 g atoms oxygen/g mol SO_2 absorbed

[b] mol $CaCO_3$ fed per mol SO_2 fed to scrubber

The physical properties of the slurry that was oxidized in the presence of adipic acid were not noticeably different from those of fully oxidized slurries obtained in tests without adipic acid: it settled at 2.5 cm/min and filtered to 76 percent solids.

A comparison of the pressure drop required to obtain a given SO_2 removal in the pilot scrubber is shown in Figure 3 for single-loop systems operating with forced oxidation. It indicates that the improved mass transfer resulting from adipic acid addition may permit a given SO_2 removal to be attained with substantially lower pressure drop than that required without the additive.

The only adverse effect of adipic acid noted at IERL-RTP was a disagreeable odor arising from the open tanks. It was present during all tests with this additive and its cause is unknown. The adipic acid feed was odorless.

The tests with sodium sulfate addition were also made in the scrubbing configuration of Figure 1 replacing the makeup water with a solution containing 66 lb Na_2SO_4/100 gal.* (79 g/l). The TCA tower was operated at L/G = 80 gal/10^6cf (10.7 liters/m^3), 3000 ppm inlet SO_2, and 6 percent oxygen. Chloride was fed to the system as HCl gas to maintain an average of 13,000 ppm Cl in the scrubbing liquor.

The principal objective of the tests was to assess the effect of sodium sulfate addition on the scaling potential. When $CaCO_3$ dissolves in the tower, the gypsum saturation differential across it is increased by the presence of high sulfate concentrations in the scrubbing liquor.

The results of eight consecutive 90-hr runs in this mode overaged 1.13% relative sulfate saturation in the scrubber feed liquor using 12 min. total EHT residence time, including 5.6 min. in the oxidizer. With 15,000 ppm average sodium concentration in the scrubbing liquor, sulfate scale was consistently observed on the bottom TCA grid when the SO_2 make-per-pass exceeded 8 m mol/l. At lower make-per-pass (e.g., with 2500 ppm inlet SO_2) scaling did not occur.

SO_2 removal efficiencies obtained in the IERL-RTP pilot plant were as good as, or better than, those obtained with fresh water makeup. As indicated by the comparison in Table 2, higher SO_2 removal and limestone utilization resulted when the fresh water was replaced by sodium sulfate solution. When compared to the results obtained without forced oxidation (and fresh water makeup) the improvement in SO_2 removal was about 10 percent.

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*The makeup solution was saturated with gypsum by adding oxidized sludge.

TABLE 2. Effect of Sodium Sulfate on TCA Scrubber Performance at Constant Limestone Feed Stoichiometry

	Fresh Water Makeup	Na_2SO_4 Solution Makeup
SO_2 removal, %	86	88
Scrubber ΔP, cm H_2O	22	18
Limestone utilization, mol %	84	91
Cl^- conc., ppm	17,000	19,000
Na^+ conc., ppm	0	26,000
Scrubber feed pH	6.0	6.3
Scrubber effluent pH	5.0	5.4
Oxidation, mol %	99	99
Settling rate, cm/min	2.5	3.0
Filter cake solids, %	73	85
Limestone feed stoichiometry	1.40	1.36

Sludge oxidized in this mode was of excellent quality, consistently filtering to 85 percent solids. It contained about 2 mg of sodium per gram of dry solid after washing with acetone. About 0.3 mg of this sodium was non-leachable by water washing.

An overall oxidation rate of 1×10^{-3} g mol/l (min) could be maintained in the pilot plant oxidizer at pH 7 when operating with 13,000 ppm Cl^- and 15,000 ppm Na^+ in the scrubbing liquor.

The tentative conclusion from these results is that an FGD system can be operated with full replacement of the makeup water by cooling tower blowdown when the scrubber is designed for an SO_2 make-per-pass below 8 m mol/l. The combination of adipic acid, forced oxidation, and water treatment appears feasible for spray towers even in high-sulfur coal situations to maximize water reuse and eliminate the uncontrolled discharge of soluble salts.

Adipic acid has clear merits as an additive for improving SO_2 removal efficiencies of scrubbers using forced oxidation.

FUTURE PLANS

A limestone "type and grind" test program will commence at RTP next month. Four different limestone types, selected by Bechtel to give a wide range of expected reactivities, will be tested at two grinds--75 percent less than 200 mesh (coarse) and 90 percent less than 325 mesh (fine). Eleven tests are planned for the four limestones over a 22 week period to compare the limestone stoichiometries required for a given SO_2 removal efficiency in the IERL-RTP scrubber. Laboratory characterization of the physical and chemical properties of the limestones will be made elsewhere in detail.

The prospects for improving sludge quality without forced oxidation will be evaluated in the IERL-RTP pilot plant according to a test program devised by Radian Corporation ("Development of a Mathematical Basis for Relating Sludge Properties to FGD-Scrubber Operating Variables," J. L. Phillips et al., EPA-600/7-78-072, NTIS No. PB 281582/AS). This test program will use a a model based on $CaSO_3 \cdot 1/2\ H_2O$ crystallization kinetics to select and evaluate scrubber modifications that might lead to the growth of larger crystals of calcium sulfite. By increasing the crystal size, it is expected that a faster settling and more filterable sludge may be produced. These tests will be undertaken after the type and grind study.

Another additive, sodium thiosulfate, will be evaluated in the lime scrubber as a possible oxidation inhibitor. Thiosulfate has been tentatively identified by Radian as an impurity in carbide lime that may be responsible for the low oxidations observed at LG&E. If small amounts are effective in reducing oxidation when added to commercial lime, the propects for unsaturated operation should be improved.

One of the two scrubbers at IERL-RTP will be converted to a sodium-based dual alkali system later this year. This scrubber will be used to provide experimental support for EPA's full-scale dual alkali demonstration facility at LG&E's Cane Run Station.

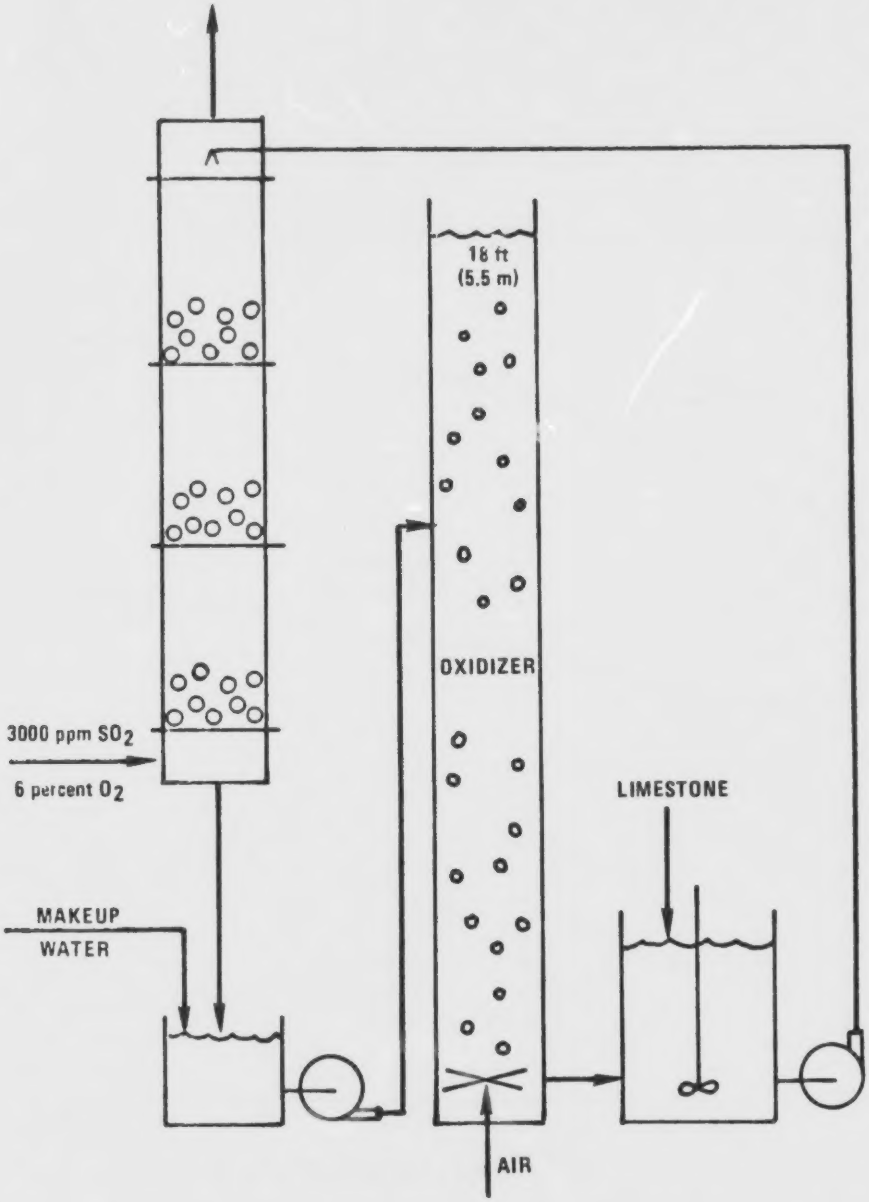
18 ft
(5.5 m)
OXIDIZER
3000 ppm SO_2
6 percent O_2
LIMESTONE
MAKEUP
WATER
AIR

Figure 1. IERL-RTP scrubber configuration for adipic acid and Na_2SO_4 tests.

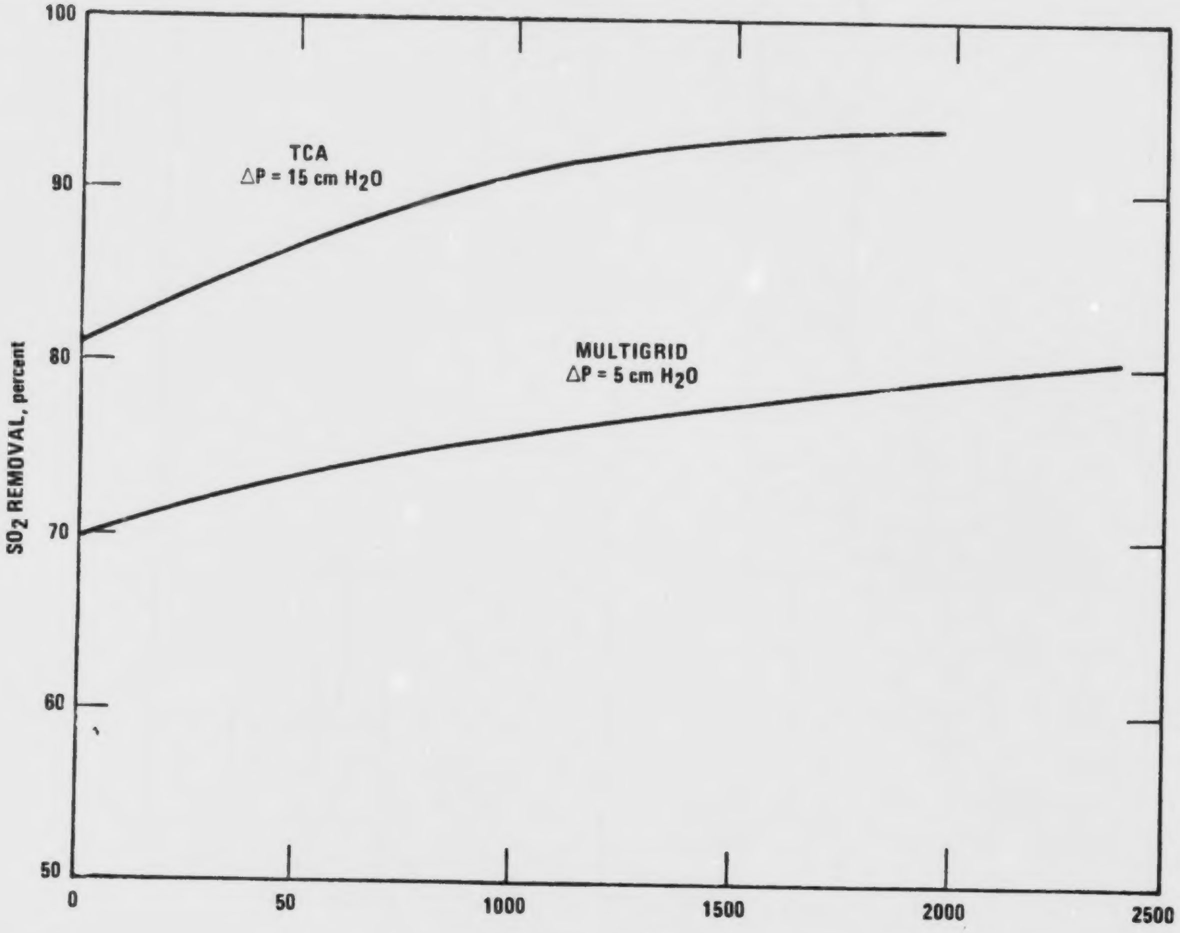

Figure 2. Effect of adipic acid on SO_2 absorption in TCA and multigrid scrubbers at L/G = 65 gal/Mcf (8.7 l/m^3).

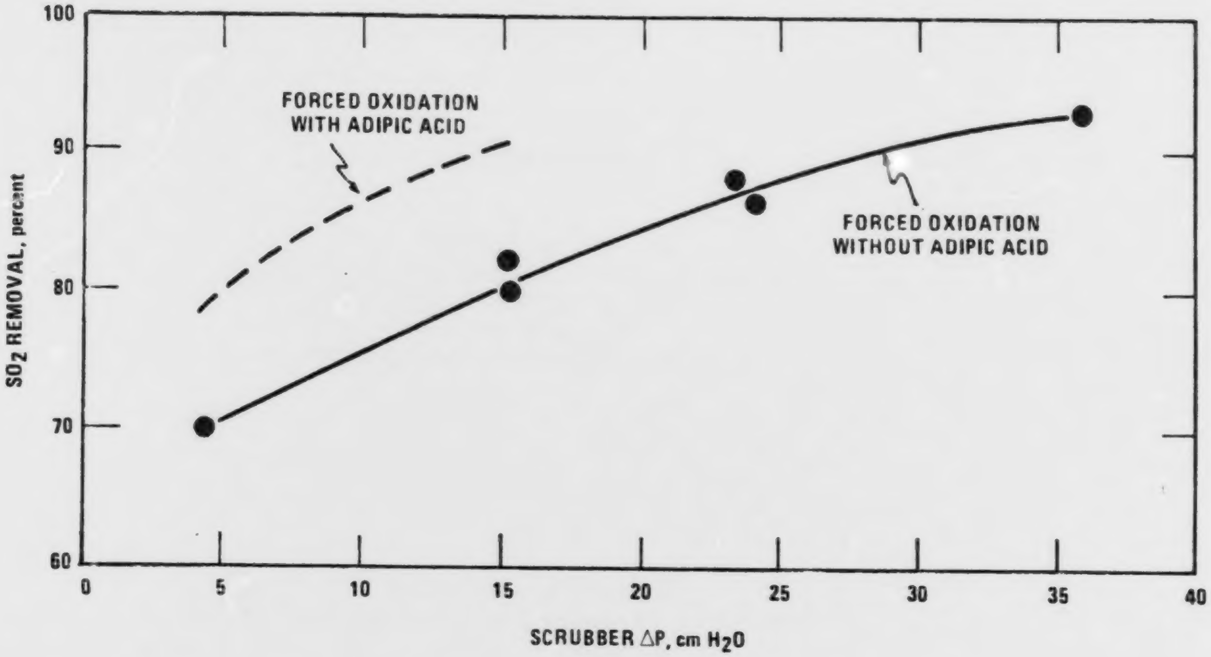

Figure 3. Effect of adipic acid on the pressure drop required to attain a given SO_2 removal efficiency in the IERL-RTP limestone scrubber.

RESULTS OF LIME AND LIMESTONE TESTING
WITH FORCED OXIDATION
AT THE EPA ALKALI SCRUBBING TEST FACILITY

- SECOND REPORT -

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at the

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August 29, 1978

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Section 1

INTRODUCTION

In lime and limestone wet-scrubbing systems for removing SO_2 and particulate from coal-fired boiler flue gas, disposal of the waste solids product has been a major problem both technically and economically. This report addresses the results of testing at the EPA Alkali Scrubbing Test Facility to develop commercially feasible forced oxidation procedures for reducing the volume and improving the disposal characteristics of the waste solids product.

The waste solids consist primarily of calcium sulfite, calcium sulfate (gypsum), and fly ash. The relative amounts of sulfite and sulfate depend on the degree of oxidation in the scrubbing system. In most medium-to-high sulfur coal applications, natural oxidation of sulfite to sulfate in the scrubber system amounts to only 10 to 30 percent and calcium sulfite is the predominant material in the waste sludge.

Calcium sulfite wastes present a serious disposal problem because of the difficulty of dewatering. The slurry can be dewatered only to about 50 to 60 percent solids, producing an unstable, thixotropic material unsuitable for landfill. Where space is available, ponding of the untreated sulfite sludge has been practiced. But the pond area may be impossible to reclaim, and in many locations sufficient space is not available.

Three procedures have been considered for converting the sulfite wastes to material suitable for landfill:

- Commercial fixation with additives
- Blending sulfite sludge with fly ash
- Forced oxidation of the calcium sulfite to a more tractable calcium sulfate (gypsum), which is easily dewatered to greater than 80 percent solids

Of these 3 procedures, preliminary economic evaluations by TVA1) have shown that forced oxidation is the most economical method to upgrade pond disposal to landfill. Cost information will be presented by TVA in a separate paper. Furthermore, because of the superior dewatering properties, forced oxidation results in a smaller volume of waste solids.

In Japan, where natural gypsum is not available, forced oxidation in scrubber systems has been employed extensively to produce a high-quality gypsum raw material for the cement and wallboard industries. In the United States, scrubber gypsum may be unable to compete extensively with the widely available natural gypsum. Thus, the incentive in the United States has been to develop simplified forced-oxidation procedures directed only toward improving waste solids handling and disposal properties. As a disposal material, the gypsum sludge can have high fly ash content; moreover, the oxidation reaction need be carried only to about 95 percent completion.

Beginning in 1976, studies conducted by EPA with the 0.1 MW pilot plant at the Industrial Environmental Research Laboratory located at Research Triangle Park, North Carolina (IERL-RTP)2), have shown that calcium sulfite can be readily oxidized to gypsum by simple air/slurry contact in the hold tank of the scrubber recirculation loop. Although the rate of oxidation reaches a maximum at a pH of 4.5 and then declines at higher pH, it was found that oxidation could be accomplished at a practical rate up to a pH of about 6.0.

Based on the findings at the IERL-RTP pilot plant, a program was set up at the Shawnee Test Facility located at the TVA Shawnee Steam Plant near Paducah, Kentucky, to develop procedures for forced oxidation. Forced oxidation testing was initiated in January 1977 on the 10 MW EPA prototype scrubbers and has continued since as the major part of the Shawnee Advanced Test Program. Results of forced oxidation testing at the Shawnee facility from January through September 1977 were reported at the FGD Symposium in Hollywood, Florida in November 1977.3) This paper is an update on that report. It includes the results of forced oxidation testing through June 1978.

Systems successfully demonstrated during this period are:

- Forced oxidation in the first of two scrubber loops using lime slurry, limestone slurry, and limestone slurry with added magnesium oxide
- Forced oxidation within a single scrubber loop using limestone slurry
- Forced oxidation of a scrubber bleed stream using limestone slurry with added magnesium oxide

THE TEST FACILITY

There are two scrubber systems operating at the EPA sponsored Shawnee Test Facility, each with its own independent slurry handling facilities. Both systems were tested with forced oxidation. The systems have the following scrubbers:

- A venturi followed by a spray tower (venturi/spray tower)
  (35,000 acfm capacity @ 300°F)
- A Turbulent Contact Absorber (TCA)
  (30,000 acfm capacity @ 300°F)

The scrubbers receive flue gas from TVA Shawnee coal-fired boiler No. 10. The boiler normally burns a high-to-medium sulfur bituminous coal producing SO_2 concentrations of 1500 to 4500 ppm. Flue gas can be taken from either side of the boiler No. 10 particulate removal equipment, allowing testing with high fly ash loadings (3 to 6 grains/scf dry) or low loadings (0.04 to 0.6 grains/scf dry). Chlorides from the flue gas concentrate in the scrubber slurry liquor over a range of 1000 to 7000 ppm depending on the tightness of the scrubber water balance and the chloride concentration in the coal burned.

The Shawnee Test Facility has been operating since March 1972. Bechtel National, Inc. of San Francisco is the major contractor and test director; TVA is the constructor and test facility operator. The inital test program lasted through October 1974(4) with the major emphasis on demonstrating reliable operation. The forced oxidation tests are a part of an advanced test program that is scheduled to continue through December 1979. Earlier results of the advanced test program are reported elsewhere.(5,6,7)

The Advanced Test Program schedule for the period covered in this report is shown in Figure 1. As can be seen, testing with forced oxidation has constituted the major effort during this period.

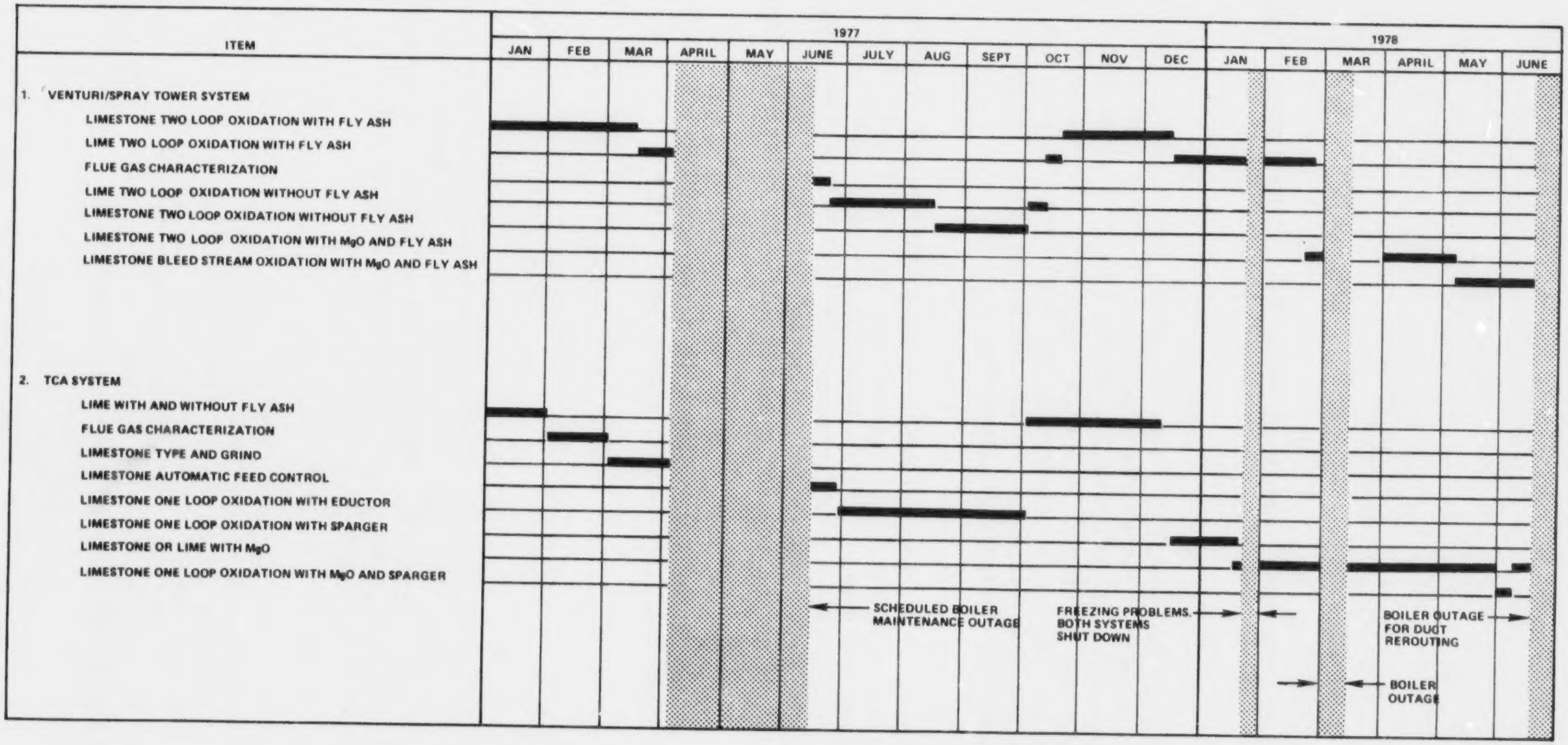


FIGURE 1. SHAWNEE ADVANCED PROGRAM TEST SCHEDULE SINCE JANUARY 1977

Section 2

FORCED OXIDATION WITH TWO SCRUBBER LOOPS ON THE VENTURI/SPRAY TOWER SYSTEM

Forced oxidation with two scrubber loops in series has been successfully demonstrated in the venturi/spray tower system with three alkali types: limestone, lime, and limestone with added magnesium oxide. In this arrangement, the flue gas passes through two scrubbers in series, each with its own hold tank and slurry recirculation loop. The first loop is operated at a relatively low pH to provide favorable conditions for forced oxidation while the second loop is operated at a higher pH for good SO_2 removal.

Commercialization of a two-loop scrubbing system with forced oxidation is both feasible and desirable in situations where land is unavailable for ponding and the higher SO_2 removal inherent with a two-loop system is required. Scrubbers in series have already been installed commercially with the first scrubber designed primarily for particulate removal and the second primarily for SO_2 removal. The addition of forced oxidation to such a system would be relatively uncomplicated.

Testing of the venturi/spray tower system in a two-scrubber-loop configuration with forced oxidation has been ongoing since January 1977.

SYSTEM DESCRIPTION

The venturi/spray tower system was modified for two-loop scrubber operation with forced oxidation as shown in Figure 2. To separate the venturi and spray tower scrubber loops, a catch funnel was installed beneath the bottom spray header of the spray tower. To eliminate slurry entrainment through the catch funnel, the bottom spray header was turned upward.

The hold tank in the first scrubber (venturi) recirculation loop was used as the oxidation tank. The arrangement of this tank is shown in Figure 3. The tank was 8 ft in diameter and could be operated at 10, 14, or 18-ft slurry levels. In early tests the tank contained an air sparger ring made of straight 3-inch 316L SS pipe pieces welded into an octagon approximately 4 ft in diameter. It was located 6 inches from the bottom of the tank. Sparger rings had either 130 1/8-inch diameter holes or 40 1/4-inch diameter holes pointing downward. The

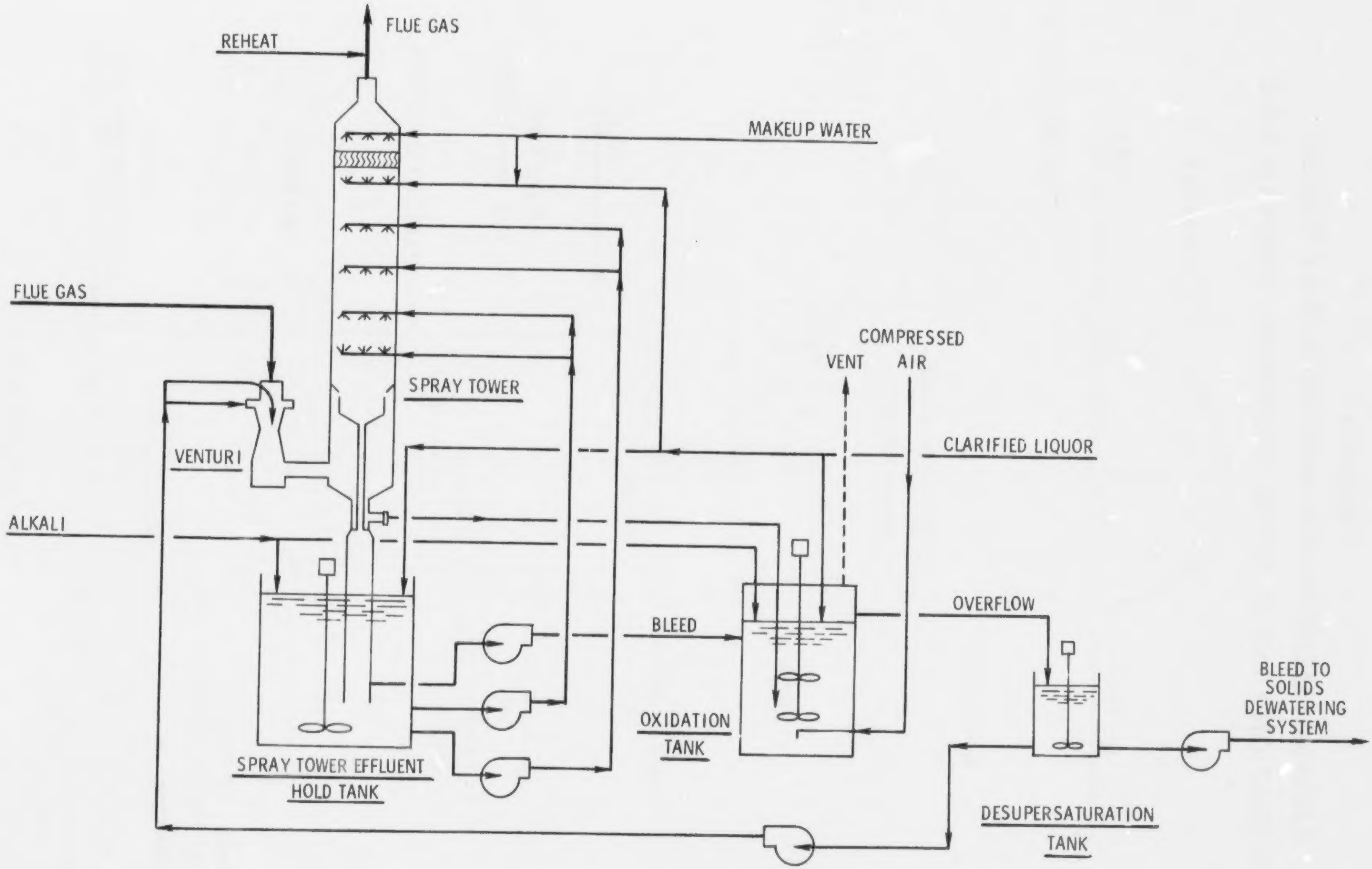
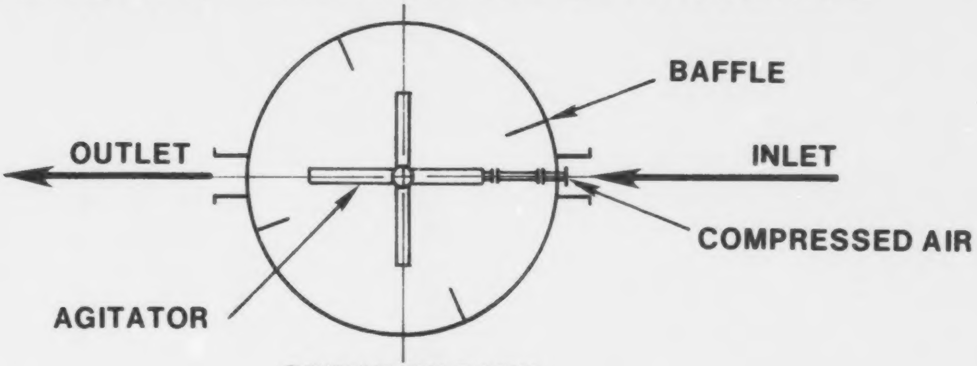
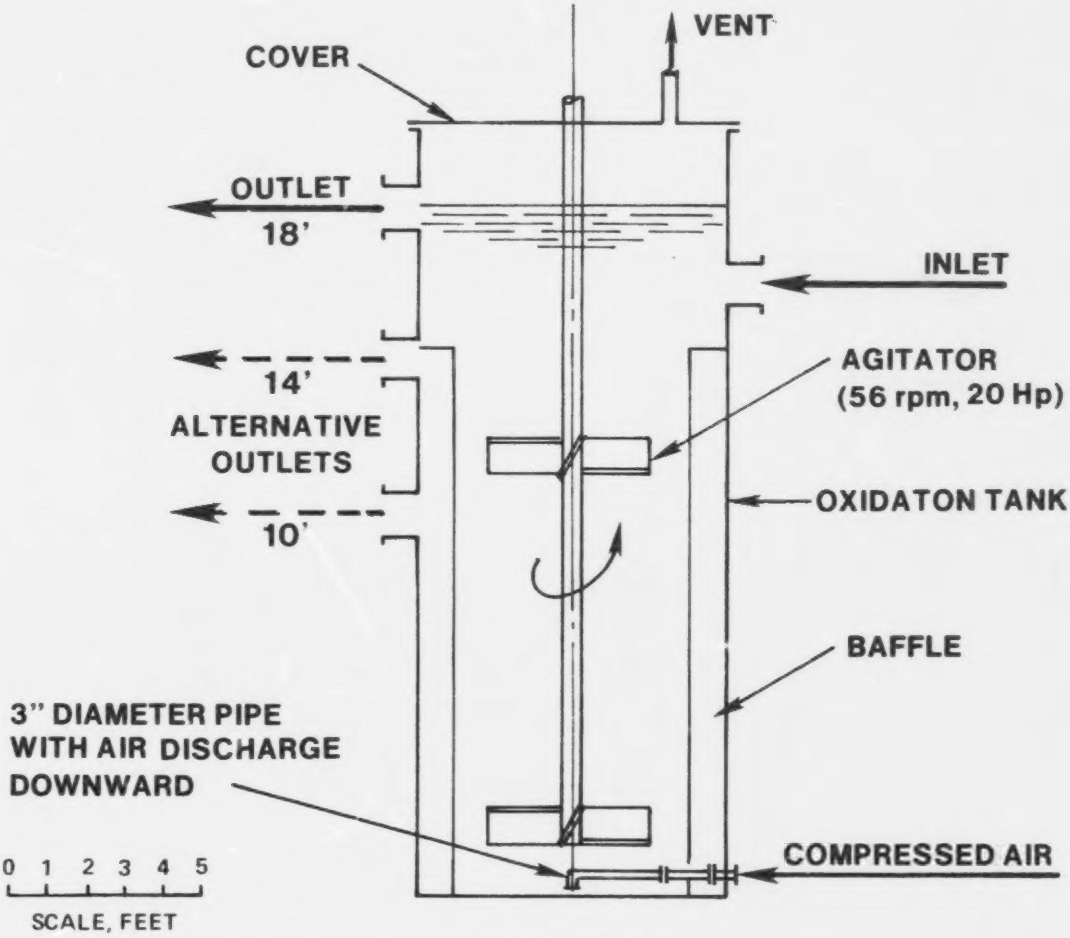

REHEAT
FLUE GAS
MAKEUP WATER
FLUE GAS
COMPRESSED
VENT
AIR
SPRAY TOWER
VENTURI
CLARIFIED LIQUOR
ALKALI
OVERFLOW
BLEED
BLEED TO
SOLIDS
DEWATERING
SYSTEM
OXIDATION
TANK
SPRAY TOWER EFFLUENT
HOLD TANK
DESUPERSATURATION
TANK

FIGURE 2. FLOW DIAGRAM FOR TWO-LOOP OXIDATION IN THE VENTURI/SPRAY TOWER SYSTEM

FIGURE 3
ARRANGEMENT OF THE VENTURI/SPRAY TOWER OXIDATION TANK WITH AIR DISCHARGE THROUGH PIPE


BAFFLE
OUTLET
INLET
COMPRESSED AIR
AGITATOR

OXIDATION TANK
PLAN VIEW


VENT
COVER
OUTLET
18'
INLET
14'
ALTERNATIVE OUTLETS
10'
AGITATOR (56 rpm, 20 Hp)
OXIDATON TANK
BAFFLE
3" DIAMETER PIPE WITH AIR DISCHARGE DOWNWARD
COMPRESSED AIR
0 1 2 3 4 5
SCALE, FEET

ELEVATION VIEW

sparger ring was fed with compressed air to which sufficient water was added to assure humidification. In more recent tests the sparger ring was replaced by a 3-inch diameter pipe with an open elbow discharging air downward at the center of the tank about 3 inches from the tank bottom.

The oxidation tank had an agitator with two axial flow turbines, both pumping downward. Each turbine was 52 inches in diameter and contained 4 blades. The bottom turbine was 10 inches above the air sparger. The agitator rotated at 56 rpm and was rated at 17 brake Hp.

A 10-ft diameter desupersaturation tank, operating at a 5-ft slurry level, followed the oxidation tank to provide time for gypsum precipitation and to provide air-free pump suction.

Provision was made to add alkali to either scrubber loop. Clarified liquor from the dewatering system could be returned to either scrubber loop or to the mist eliminator wash circuit.

SUMMARY OF PREVIOUSLY REPORTED TEST RESULTS

Forced oxidation test results with two scrubber loops conducted from January through mid-September 1977 with lime and limestone slurry have been previously reported.3) These early tests were conducted at 25,000 acfm flue gas rate (300^{o}F) which corresponds to a superficial gas velocity in the spray tower of 6.7 ft/sec. A flue gas rate lower than the maximum possible in the system (35,000 acfm) was chosen to assure that high SO_2 removal (greater than 80 percent) could be achieved. Slurry recirculation rates of 600 gpm in the venturi loop and 1400 gpm in the spray tower loop were used. Each run averaged about 5 to 6 days which was judged to be sufficient time to reach kinetic equilibrium and to allow adequate run data to be gathered.

Key results from these earlier tests were as follows:

- Oxidation of sulfite solids to gypsum of 90 percent or better dramatically improved the dewatering and handling characteristics of the waste solids.
- Slurry oxidation of better than 96 percent in the first of two independent scrubbing loops was demonstrated with simple air sparging through a sparger ring in an open tank with the configuration shown in Figure 3.
- Conditions under which near complete oxidation was demonstrated were an oxidation tank pH range of 4.5 to 5.5, an air stoichiometric ratio

of at least 1.5 atoms O/mole of SO_2 absorbed, and an oxidation tank level of at least 14 feet.

- Slurries with high or low fly ash loadings oxidized equally well.
- A slurry solids concentration of 7 percent or higher in the spray tower was required to prevent calcium sulfite scaling and to maintain good SO_2 removal.
- For pH control, it was necessary to add lime to both scrubber loops. With limestone, addition to the spray tower loop was sufficient.

UPDATE ON TWO-SCRUBBER-LOOP TEST RESULTS WITH LIME SLURRY

Since the last report, 7 runs have been made with lime slurry in a two-scrubber-loop configuration with forced oxidation. Results of the tests are summarized in Table 1. These tests further demonstrated the feasibility of a two-scrubber-loop forced-oxidation system and contributed more information for commercial design. In all these runs, the filter cake solids concentration was at least 80 percent and usually above 85 percent. The following describes design factors developed during these tests.

Air Sparger - Previous tests were made with air dispersed into the oxidation tank through a sparger ring containing either 1/8-inch or 1/4-inch holes. Some plugging and erosion of the holes were experienced with these rings. Starting with Run 861-1A the sparger ring was replaced with a 3-inch pipe with an elbow on the end to direct the air downward at the center of the tank about 3 inches above the tank bottom and 1 foot below the bottom agitator blade. Oxidation efficiency was as good with this 3-inch air pipe as with the spargers. For example, in Run 861-1A with the air pipe, 98 percent sulfite oxidation was achieved at an air stoichiometry of 1.5 atoms O/mole SO_2 absorbed. Based on the success with the open 3-inch pipe, it was concluded that the agitator plays a primary role in dispersing the air. The agitator used in these tests had two sets of axial flow turbine blades pumping downward and operated at a fixed speed of 56 rpm with 17 brake Hp. A variable speed agitator, presently on order, will be installed so that information can be obtained on agitator power versus oxidation efficiency.

Slurry solids concentration - Virtually all the fly ash is captured in the slurry in the venturi loop so the spray tower slurry is essentially fly ash free. The venturi loop is normally controlled at a higher slurry solids concentration than the spray tower loop to compensate for the fly ash.

Slurry solids concentration in the venturi scrubber loop was controlled at 15 weight percent in a majority of the lime runs. Slurry solids concentration in the spray tower varied from 6 percent to almost 20 percent depending on whether the

Table 1

RESULTS OF FORCED OXIDATION TESTS WITH TWO SCRUBBER LOOPS ON THE VENTURI/SPRAY TOWER SYSTEM USING LIME SLURRY

Major Test Conditions	861-1A	862-1A	863-1A	864-1A	865-1A	866-1A	867-1A
Fly ash loading	Low	High	High	High	High	High	High
Flue gas rate, acfm @ 300°F	25,000	35,000	Variable	35,000	35,000	35,000	35,000
Slurry rate to venturi, gpm	600	600	600	600	600	600	600
Slurry rate to spray tower, gpm	1400	1400	1600	1600	1600	1600	1600
Venturi percent solids recirculated (controlled)	15	15	15	15	15	15	15
Residence times, min: Oxidation tank	11.3	11.3	11.3	11.3	6.3	6.3	8.8
Desupersaturation tank	4.7	4.7	4.7	4.7	4.7	4.7	4.7
Spray tower EHT	18	18	14.7	14.7	14.7	14.7	14.7
Venturi inlet (oxidation tank) pH (controlled)	5.5	5.5	5.5	5.5	5.5	5.5	5.5
Spray tower inlet pH (controlled)	8.0	8.0	7.8	7.8	7.8	7.8	7.8
Venturi pressure drop, in. H_2O	9	9	≦9	9	9	9	9
Oxidation tank level, ft	18	18	18	18	10	10	14
Air rate to oxidation tank, scfm(1)	150	210	210	210	290/210	350	210
Clarified liquor returned to (2)	S.T.	Vent	Vent & S.T.	Vent & S.T.	Vent & S.T.	Vent & S.T.	Vent & S.T.
Selected Results							
Onstream hours	117	162	779	115	254	159	137
Percent SO_2 removal	92	85	88	94	89	94	89
Inlet SO_2 concentration, ppm	2700	2650	2950	2250	2300	1700/2400	2300
Spray tower percent solids recirculated	7.3	17.3	10.4	9.5	10.1	9.9	11.8
Spray tower lime stoichiometric ratio	1.16	1.11	1.11	1.13	1.10	1.15	1.12
Spray tower inlet liquor gypsum saturation, %	90	85	100	100	95	90	85
Spray tower sulfite oxidation, %	19	18	21	24	26	26	20
Overall sulfite oxidation, %	98	97	97	98	89	98/81	98
Overall lime utilization, %	99	98	98	99	98	98	99
Venturi inlet liquor gypsum saturation, %	95	105	105	100	100	95	90
Venturi inlet liquor sulfite concentration, ppm	35	30	35	25	65	40	40
Air stoichiometry, atoms O/mole SO_2 absorbed	1.50	1.65	1.40-2.75	1.75	2.10	3.85-2.70	1.80
Filter cake solids, wt%(3)	86	86	85	85	80	81	86
Mist eliminator restriction, %(4)	1	1.5	0	-	2	3	3

Notes:
1) Air discharged through 3-inch diameter pipe with discharge downward through an open elbow at center of oxidation tank about 3-inches from tank bottom.
2) Spray tower loop (effluent hold tank) or venturi loop (oxidation tank).
3) Clarifier and filter in series used for solids dewatering in all runs.
4) Intermittent mist eliminator bottom wash with makeup water at 1.5 gpm/ft^2 for 6 minutes every 4 hours. Sequential top wash with makeup water with one of 6 nozzles on at 0.53 gpm/ft^2 for 4 minutes every 80 minutes.

clarifed liquor from the dewatering system was returned to the venturi loop or the spray tower loop. Beginning with Run 863-1A, the returning clarified liquor was split between the two scrubber loops to control the spray tower slurry solids concentration at about 10 percent. This solids level was a compromise between low solids (below about 7 percent) where scaling and a drop in SO_2 removal is experienced and high solids (above about 15 percent) where it becomes more difficult to keep the mist eliminator clean.

Flue Gas Rate - Earlier runs were made at a reduced flue gas flow rate of 25,000 acfm (at 300°F) because it was assumed that forced oxidation would reduce SO_2 removal efficiency in the venturi loop at lower pH to the extent that an overall SO_2 removal efficiency of at least 80 percent could not be achieved. This proved not to be the case. Beginning with Run 862-1A, the flue gas flow rate was increased to maximum achievable of 35,000 acfm (at 300°F) which corresponds to a spray tower superficial velocity of 9.4 ft/sec. In this run and subsequent runs at 35,000 acfm, SO_2 removal averaged about 85 percent at 2000 to 3000 ppm inlet SO_2 concentration.

Spray tower slurry liquor pH - In earlier runs the spray tower inlet slurry liquor pH was controlled at 8.0. In some runs, especially those at low slurry solids concentration as mentioned previously, this level of inlet pH resulted in an outlet pH approaching 6, causing sulfite scaling.

While the pH drop across the spray tower depends on the SO_2 removal and the inlet SO_2 concentration, it has been generally observed that sulfite scaling does not occur if the spray tower outlet pH stays below about 5.5 Therefore, during Run 863-1A, the inlet pH was adjusted downward slightly to 7.8. In this run and in subsequent runs, small patches of scale were observed to appear and disappear in a cyclic manner. This cyclic appearance of scale did not interfere with scrubber operation. No effect on SO_2 removal was discernable as a result of the slight adjustment in inlet pH.

This slight adjustment in pH is significant in that it demonstrates the need for good pH control in commercial installations and it demonstrates one of several operating adjustments that can be made to eliminate a scaling problem.

Slurry Level in Oxidation Tank - In Runs 864-1A through 867-1A, the effect of oxidation tank slurry level (and consequently air and slurry residence times) was explored. In these tests, air was discharged into the bottom of the oxidation tank through an open 3-inch pipe as previously described (see Figure 3). Major test conditions are listed in Table 1. All runs were made at an oxidation pH of 5.5. The effect of the tank level is summarized below:

Run No.	Oxidation Tank Level, ft	Air Stoichiometry, atoms O/mole SO_2 absorbed	Percent Sulfite Oxidation
864-1A	18	1.8	98
867-1A	14	1.8	98
865-1A	10	2.1	89
866-1A	10	3.8/2.7	98/81

Oxidation efficiency was high at 18 and 14-ft tank levels but dropped off at a 10-ft level. In Run 866-1A, high oxidation efficiency was achieved at a 10-ft tank level by increasing air stoichiometry. Part of Run 866-1A was made at lower air stoichiometry with subsequent loss in oxidation efficiency.

These runs demonstrated that 98 percent sulfite oxidation can be achieved at 14 to 18-ft tank levels at an air stoichiometry of 1.8 atoms O/mole SO_2 absorbed. At a 10-ft tank level an air stoichiometry approaching 3.8 is required.

It must be pointed out that the oxidation at a 10-ft tank level is not directly comparable with those at 14 and 18 feet because the top turbine of the agitator is located at the 11-ft level. In 10-ft slurry level tests, the top turbine is not in contact with the slurry and a different agitation pattern results.

Filter cake solids concentration during these tests was about 85 percent when the oxidation efficiency was 98 percent. In test periods when oxidation efficiency dropped below 90 percent, the filter cake solids concentration tended toward a lower range of 80 percent.

Lime Reliability Run - From mid-December 1977 through mid-January 1978, Run 863-1A, a one-month lime-slurry reliability run, was made with the venturi/spray tower system in a two-scrubber-loop configuration with forced oxidation in the venturi scrubber loop. Onstream operation for this run totaled 779 hours (32 days). The run was designed to demonstrate operating reliability of the scrubber system with respect to scaling and plugging and to determine if the EPA New Source Performance Standards for SO_2 and particulate emissions could be met.

To simulate variable boiler load, the flue gas flow rate was varied between 18,000 and 35,000 acfm (4.8 and 9.4 ft/sec spray tower superficial gas velocity) as the boiler load varied between 100 and 150 MW. Flue gas with high fly ash loading was used. The venturi plug was fixed at a position to give 9 inches H_2O

pressure drop across the venturi at full 35,000 acfm flue gas flow rate. The actual venturi pressure drop ranged from 2 to 9 inches H_2O. The slurry recirculation rates to the venturi and spray tower were held constant at 600 and 1600 gpm, respectively. The venturi inlet pH was controlled at 5.5. The oxidation tank level was 18 ft and the oxidation air flow rate was 210 scfm discharged through a 3-inch pipe. As previously discussed, the spray tower slurry inlet pH was adjusted downward from 8.0 to 7.8 to eliminate an observed sulfite scale buildup.

During the run, the scrubber was shut down a total of 57 hours; 46 hours were due to boiler outages, 7-1/2 hours were for scheduled scrubber inspections, and 3-1/2 hours were unscheduled downtime. This resulted in a scrubber availability of 99.6 percent, excluding the interruptions due to boiler outage and the scheduled inspections. The unscheduled downtime included 2 hours for mist eliminator cleaning and 1-1/2 hours for air compressor repair.

Average SO_2 removal for the entire run was 88 percent at 2950 ppm average inlet SO_2 concentration. This corresponds to an average emission of 0.9 lb SO_2/MM Btu, well within the EPA standard of 1.2 lb SO_2/MM Btu. However, due to unusually wide fluctuations in inlet SO_2 concentration and slow system response time, the SO_2 emissions at times exceeded the EPA standard for periods greater than the three hours allowed by EPA regulations.

The fluctuations in inlet SO_2 concentration, ranging up to 4700 ppm, resulted from the wide variety of coals being burned during the 1977-78 coal strike. Normally, inlet SO_2 concentration ranges between 2000 and 3000 ppm. These high SO_2 concentrations were beyond the capacity of the venturi/spray tower system to remove with its limited slurry recirculation rates (liquid-to-gas ratios of 57 and 21 gal/Macf at 35,000 acfm full gas flow rate in the venturi and spray tower, respectively).

Average particulate loading was 0.046 grain/dry scf corresponding to an average emission of 0.09 lb particulate/MM Btu (assuming 30 percent boiler excess air). Although the EPA standard of 0.10 lb particulate/MM Btu was not exceeded on the average, a few measurements exceeded this value.

Sulfite oxidation averaged 97 percent during the run with the air stoichiometric ratio varying between 1.4 and 2.8 atoms O/mole SO_2 absorbed. The filter cake was excellent throughout the run with solids concentration averaging 85 percent. Lime utilization was 90 percent in the spray tower and 98 percent overall, reflecting the high utilization to be expected in a two-scrubber-loop system.

At the first scheduled inspection after 160 operating hours, the mist eliminator was found to be 15 percent restricted by solids. After a review of the history of the mist eliminator exposure, the restriction was attributed to excess calcium carbonate from the previous limestone run (limestone stoichmetric ratio of 1.65

in the spray tower) and a failure to activate the intermittent underwash for the first eight hours of the reliability run. At the beginning of the reliability run, the mist eliminator underwash had been changed from continuous with diluted clarified liquor (needed for the limestone run conditions) to intermittent with makeup water (1.5 gpm/ft^2 for 6 minutes every 4 hours - satisfactory for lime runs). The mist eliminator was cleaned and the run was continued. This mishap was frustrating in that it broke a record of 4183 hours of operation under widely varying conditions without cleaning the mist eliminator.

At subsequent inspections at 399 operating hours and at the end of the run, the mist eliminator was entirely clean.

In summary, the operating reliability of the venturi/spray tower system in a two-scrubber-loop configuration with forced oxidation in lime slurry service has been demonstrated with a system availability of 99.6 percent. However, under the conditions selected, the system was unable to continually meet EPA New Source Performance Standards for SO_2 and particulate emissions even though the average emissions for the run met the standards.

UPDATE ON TWO-SCRUBBER-LOOP TEST RESULTS WITH LIMESTONE SLURRY

Since the last report, 6 runs have been made with limestone slurry in a two-scrubber-loop configuration with forced oxidation. Results of the tests are summarized in Table 2. In these tests, several operating problems were solved and operational reliability was established. Filter cake solids concentration stayed consistantly above 85 percent throughout the tests. The following discussion highlights the new information developed from the recent runs.

Flue Gas Rate - As with lime testing, earlier limestone runs were made at a reduced flue gas flow rate under the assumption that, with forced oxidation, high SO_2 removal (in the range of 85 percent) could not be achieved at full gas rate. Beginning with Run 815-1A, the flue gas flow rate was increased from the reduced rate of 25,000 acfm (at 300°F) to the maximum rate of 35,000 acfm. SO_2 removal for this run was 86 percent under the test conditions listed in Table 2, which is about 5 percentage points below SO_2 removal achieved on an identical run at the lower flue gas flow rate. All subsequent runs were made at the higher flue gas flow rate.

System Control - In the limestone tests with two scrubber loops, the control philosophy was to hold the venturi inlet pH (oxidation tank pH) at 5.5 by adjusting the limestone slurry feed rate to the spray tower effluent hold tank. Control in this manner proved to be difficult and wide fluctuations were experienced in both pH and limestone stoichiometry. For example, in Run 815-1A, the venturi inlet pH varied between 4.9 and 6.3 with corresponding fluctuations in the limestone stoichiometric ratios of 1.1 to 1.9 in the venturi loop and 1.2 to 2.7 in the spray tower loop.

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Table 2

RESULTS OF FORCED OXIDATION TESTS WITH TWO SCRUBBER LOOPS ON THE VENTURI/SPRAY TOWER SYSTEM USING LIMESTONE SLURRY

Major Test Conditions	815-1A	816-1A	817-1A	818-1A	819-1A	819-1B
Fly ash loading	Low	Low	High	High	High	High
Flue gas rate, acfm @ 300^{o}F	35,000	35,000	35,000	35,000	Variable	Variable
Slurry rate to venturi, gpm	600	600	600	600	600	600
Slurry rate to spray tower, gpm	1400	1400	1400	1600	1600	1600
Venturi percent solids recirculated (controlled)	15	15	15	15	15	15
Residence times, min: Oxidation tank	8.8	11.3	11.3	11.3	11.3	11.3
Desupersaturation tank	4.7	4.7	4.7	4.7	4.7	4.7
Spray tower EHT	13.4	13.4	16.8	14.7	14.7	14.7
Venturi inlet (oxidation tank) pH (controlled)	5.5	5.5	5.5	5.5	5.5	-
Spray tower limestone stoichiometric ratio	-	-	-	-	-	1.6
Venturi pressure drop, in. H_2O	9	9	9	7.5-9	≤9	≤9
Oxidation tank level, ft	14	18	18	18	18	18
Air rate to oxidation tank, scfm	210(1)	210(1)	210(2)	210(2)	210(2)	210(2)
Clarified liquor returned to (3)	S.T	S.T.	Vent	Vent	Vent	Vent
Selected Results						
Onstream hours	306	142	188	141	840	126
Percent SO_2 removal	86	86	83	86	86	85
Inlet SO_2 concentration, ppm	2500	2350	2500	2550	2950	3000
Spray tower percent solids recirculated	8.4	7.7	8.9	9.6	10.0	9.6
Spray tower inlet pH	5.85	5.75	5.9	5.9	5.85	5.9
Spray tower limestone stoichiometric ratio	1.98	1.68	1.60	1.64	1.65	1.65
Spray tower inlet liquor gypsum saturation, %	105	105	100	100	100	110
Spray tower sulfite oxidation, %	26	27	21	19	21	19
Overall sulfite oxidation, %	96	98	97	98	98	98
Overall limestone utilization, %	67	83	82	81	81	83
Venturi inlet liquor gypsum saturation, %	105	100	105	105	100	105
Venturi inlet liquor sulfite concentration, ppm	40	25	25	25	25	25
Air stoichiometry, atoms O/mole SO_2 absorbed	1.70	1.80	1.75	1.70	1.45-2.80	1.45-2.80
Filter cake solids, wt%(4)	87	86	86	86	87	86
Mist eliminator restriction, %(5)	3	3	1	2	3	3

Notes:

1) Air discharged downward through sparger ring with 40-¼ inch diameter holes about 3 inches from tank bottom.
2) Air discharged downward through 3-inch diameter pipe with an open elbow at center of oxidation tank about 3 inches from tank bottom.
3) Spray tower loop (effluent hold tank) or venturi loop (oxidation tank).
4) Clarifier and filter in series used for solids dewatering in all runs.
5) Continuous mist eliminator bottom wash with diluted clarified liquor at 0.4 gpm/ft^2. Sequential top wash with makeup water with one of 6 nozzles on at 0.53 gpm/ft^2 for 4 minutes every 80 minutes.

In Run 815-1A, the oxidation tank level was 14 feet, which was satisfactory for forced oxidation (96 percent oxidation at an air stoichiometric ratio of 1.7 atoms O/mole SO_2 absorbed). In Run 816-1A, the fluctuation in venturi inlet pH was reduced to a range of 5.2 to 5.8 with corresponding reduction in fluctuation in limestone stoichiometry by increasing the oxidation tank level to the maximum of 18 feet and thus increasing hold tank residence time.

In Run 819-1B, the control philosophy was changed. In this run, the limestone stoichiometry in the spray tower was controlled at 1.6 moles calcium per mole SO_2 absorbed and the venturi inlet pH was allowed to vary. With direct control on the spray tower stoichiometry, the fluctuation in venturi inlet pH was 5.2 to 5.8, no greater than in the previous runs with venturi inlet pH control.

Based on these runs, control of limestone stoichiometry in the primary scrubbing loop (spray tower) is recommended over control of pH in the oxidation loop (venturi).

Mist Eliminator - In previous runs with limestone slurry and high fly ash loadings (Runs 805-1A through 808-1A), problems with mist eliminator plugging occured. In these runs, the spray tower solids concentration was maintained at 15 percent, which required that the clarified liquor from the solids dewatering system be returned to the venturi loop and which allowed only enough makeup water in the spray tower system for an intermittent mist eliminator underside wash. Such a wash was inadequate at the limestone utilizations experienced in the spray tower (60 to 70 percent) and the mist eliminator plugged within a matter of days.

Beginning with Run 817-1A, the mist eliminator was washed continuously at 0.4 gpm/ft^2 with clarified water diluted with available makeup water. Excess clarified water was returned to the venturi loop. This wash scheme (coupled with a sequential top wash - see Table 2) proved adequate and the mist eliminator no longer plugged.

The continuous wash diluted the spray tower solids concentration to about 9 percent. At 9 percent solids concentration, SO_2 removal dropped a few percentage points to 83 percent at 2500 ppm inlet SO_2 concentration.

In Run 818-1A, the slurry recirculation rate in the spray tower loop was increased from 1400 gpm to the maximum controlled rate of 1600 gpm. With this modification, SO_2 removal was increased to 86 percent at 2550 ppm inlet concentration.

Limestone Reliability Run - During November 1977, Run 819-1A, a one-month limestone slurry reliability run, was made with a two-scrubber-loop configuration on the venturi/spray tower system and with forced oxidation in the venturi loop. This run operated for a total of 840 hours (35 days). As with the lime reliabilty run, the run was designed to demonstrate operating reliability of the scrubber system and to determine if the EPA New Source Performance Standards for SO_2 and particulate emission could be met.

Flue gas and slurry flow rates were the same as with the lime reliability run. Flue gas with high fly ash loading was varied in rate between 18,000 and 35,000 acfm (at 300°F) to follow the boiler load. The venturi plug was fixed to give 9 inches H_2O pressure drop at 35,000 acfm flue gas rate. Slurry recirculation rates were held constant at 600 gpm and 1600 gpm in the venturi and spray tower loops respectively. The venturi inlet pH was controlled at 5.5 by controlling the limestone feed rate to the spray tower hold tank.* The oxidation tank was maintained with an 18-ft slurry level and an air flow rate of 210 scfm discharged through a 3-inch pipe.

During the run, the scrubber was shut down for 18 hours due to a boiler outage, 5 hours total for weekly inspections, plus 3-1/2 hours of unscheduled downtime for a total of 26-1/2 hours. Based on unscheduled downtime, the scrubber system had an availability of 99.6 percent. The unscheduled downtime included 3 hours to clean a partially plugged slurry return pipe from the venturi to the oxidation tank and 1/2 hour to clean a plugged mist eliminator nozzle.

The plugged mist eliminator nozzle was discovered after 391 hours of operation. The mist eliminator in the vicinity of the slurry nozzle was severely restricted by slurry solids (7 percent overall mist eliminator restriction). The nozzle was cleaned but the mist eliminator was not disturbed. By the end of the run (840 hours), the mist eliminator restriction had dropped to 3 percent, demonstrating that a restricted area can be self cleaning.

For the entire run, the SO_2 removal averaged 86 percent at 2950 ppm average inlet SO_2 concentration. This removal efficiency corresponds to an average emission of 1.0 lb SO_2/MM Btu which meets the EPA New Source Performance Standard of 1.2 lb SO_2/MM Btu. However, as with the lime reliability run, fluctuations to unusually high inlet SO_2 concentrations were experienced and the standard was at times exceeded for periods greater than the three hours allowed by EPA regulations.

The outlet particulate loading ranged from 0.021 to 0.063 grain/dry scf with an average of 0.042 grain/dry scf. Assuming 30 percent excess air to the boiler, the average outlet particulate loading corresponds to 0.08 lb/MM Btu which meets the EPA New Source Performance Standard of 0.1 lb/MM Btu. However, a few of the outlet particulate loading measurements exceeded the standard.

Sulfite oxidation averaged 98 percent during the run with the air stoichiometric ratio varying between 1.4 and 2.8 atoms O/mole SO_2 absorbed. The filter cake solids concentration averaged 87 percent. Overall limestone utilization was 81 percent while the spray tower limestone utilization was 61 percent, again demonstrating the advantage of a two-scrubber-loop system in achieving high alkali utilization.

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* As previously discussed, this mode of control was later changed to stoichiometric ratio control in the spray tower (Run 819-1B).

To summarize, the operating reliability of the venturi/spray tower system in a two-scrubber-loop configuration with forced oxidation in limestone slurry service has been demonstrated with a system availability of 99.6 percent. However, under the conditions selected, the system was unable to continually meet EPA New Source Performance Standards for SO_2 and particulate emissions even though the average emissions for the run met the standards.

TWO-SCRUBBER-LOOP TEST RESULTS USING LIMESTONE SLURRY WITH ADDED MAGNESIUM OXIDE

Beginning in March 1978, a series of six runs were made in which magnesium oxide was added to the spray tower hold tank along with the limestone slurry. The primary purpose of the magnesium oxide addition was to enhance SO_2 removal efficiency in the spray lower loop by increasing the sulfite ion concentration in the liquor for SO_2 scrubbing. In a two-scrubber-loop configuration as shown in Figure 2, the magnesium ion concentration in the venturi loop is higher than that in the spray tower loop because of the water loss in humidifying the flue gas in the venturi loop. But because the sulfite ion is converted into nonscrubbing sulfate ion by forced oxidation, the higher magnesium ion concentration in the venturi loop does not enhance SO_2 removal in the venturi loop. The secondary purpose of the magnesium oxide addition was to determine whether the presence of magnesium ion had an effect on oxidation efficiency.

Typical operating conditions and results of these tests are summarized in Table 3. The expected enhancement of SO_2 removal was achieved and oxidation efficiency was, if anything, improved. Thus, magnesium oxide addition is compatible with a two-scrubber-loop forced oxidation system.

SO_2 Removal - Run 820-1A was made under identical conditions to Run 818-1A (Table 2) except for the addition of magnesium oxide. Effective magnesium ion concentration* averaged 5150 ppm in the spray tower. The anticipated removal enhancement was achieved as the average SO_2 removal was 96 percent at 2250 ppm average inlet SO_2 concentration compared with 86 percent removal at 2550 inlet ppm for Run 818-1A.

The spray tower inlet slurry liquor was 100 percent saturated in gypsum and no scale was observed. This condition was typical of all the tests in the limestone/magnesium oxide, forced-oxidation test block.

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* Effective magnesium ion concentration is defined as the total magnesium ion minus that magnesium ion concentration equivalent to total chlorides. Magnesium chloride has no effect on SO_2 removal.

Table 3

RESULTS OF FORCED OXIDATION TESTS WITH TWO SCRUBBER LOOPS

ON THE VENTURI/SPRAY TOWER SYSTEM USING LIMESTONE SLURRY WITH ADDED MAGNESIUM OXIDE

Major Test Conditions	820-1A	820-1B	820-1C	821-1A	822-1A	822-1B
Fly ash loading	High	High	High	High	High	High
Flue gas rate, acfm @ 300°F	35,000	35,000	35,000	35,000	35,000	35,000
Slurry rate to venturi, gpm	600	600	600	600	600	600
Slurry rate to spray tower, gpm	1600	1600	1600	0	1600	1600(5)
Venturi percent solids recirculated (controlled)	15	15	15	15	15	15
Residence times, min: Oxidation tank	11.3	11.3	11.3	11.3	11.3	11.3
Desupersaturation tank	4.7	4.7	4.7	4.7	4.7	4.7
Spray tower EHT	14.7	14.7	14.7	-	14.7	14.7
Venturi inlet (oxidation tank) pH (controlled)	5.5	-	-	5.5	-	-
Spray tower limestone stoichiometric ratio (based on solids)	-	1.6(6)	1.6(6)	-	1.6(6)	1.6(6)
Effective Mg^{++} concentration (S.T. loop), ppm	5000	5000	5000	5000	5000	5000
Venturi pressure drop, in. H_2O	9	9	9	9	9	9
Oxidation tank level, ft	18	18	18	18	18	18
Air rate to oxidation tank, scfm(1)	210	150	0	210	210	210
Clarified liquor returned to (2)	Vent	Vent	Vent	Vent	Vent	Vent
Selected Results						
Onstream hours	462	137	134		232	85
Percent SO_2 removal	96	94	91	(7)	91	90
Inlet SO_2 concentration, ppm	2250	2500	2750		2750	2400
Spray tower percent solids recirculated	6.0	8.3	10.5		8.0	5.6
Spray tower inlet pH	6.05	5.9	5.9		5.75	5.55
Spray tower limestone stoichiometric ratio (based on total slurry)	1.16	1.28	1.52		1.55	1.21
Spray tower inlet liquor gypsum saturation, %	100	105	90		100	110
Spray tower sulfite oxidation, %	30	17	20		21	23
Effective Mg^{++} concentration (S.T. loop), ppm	5150	4985	4700		4985	4845
Overall sulfite oxidation, %	98	92	36		97	98
Overall limestone utilization, %	92	90	82		79	93
Venturi inlet liquor gypsum saturation, %	130	130	145		125	130
Venturi inlet liquor sulfite concentration, ppm	50	950	5585		735	410
Air stoichiometry, atoms O/mole SO_2 absorbed	1.70	1.10	0		1.45	1.70
Filter cake solids, wt% (3)	85	82	63		85	85
Mist eliminator restriction, %(4)	-	0	0	0	-	0

Notes:

1) Air discharged downward through 3-inch diameter pipe with an open elbow at center of oxidation tank about 3 inches from tank bottom.
2) Venturi loop (oxidation tank).
3) Clarifier and filter used for solids dewatering in all runs.
4) Continuous mist eliminator bottom wash with diluted clarified liquor at 0.4 gpm/ft^2. Sequential top wash with makeup water with one of 6 nozzles on at 0.53 gpm/ft^2 for 4 minutes every 80 minutes.
5) Spray tower turned off for 30 minutes every 8 hours to obtain SO_2 removal with venturi alone. Venturi SO_2 removal averaged 29%.
6) In runs with control by spray tower stoichmetric ratio, the venturi inlet pH averaged 5.0.
7) Run failed due to low MgO dissolution rate in spray tower effluent hold tank.

In Run 821-1A, an attempt was made to determine the SO_2 removal in the venturi by turning off the slurry recirculation to the spray tower. Unfortunately, the magnesium oxide, added to the spray tower hold tank, would not dissolve without recirculation and the run was aborted.

A second effort was more successful. Run 822-1B was an extension of 822-1A in which the spray tower slurry recirculation was turned off once a shift for only 30 minutes. This short time period did not upset the system balance. SO_2 removal in the venturi loop was found to be 29 percent which is typical of removal efficiency with limestone slurry in the absence of magnesium ion. Thus, it has been demonstrated that magnesium ion does not enhance SO_2 removal in a scrubber loop with forced oxidation.

Run 822-1A was made in an effort to improve removal efficiency by minor changes in piping configuration to locate makeup and bleed streams at their optimum locations in the venturi slurry recirculation loop. Referring to Figure 2, the bleed from the spray tower loop was sent to the desupersaturation tank instead of the oxidation tank as shown. Also, the bleed to the solids dewatering system was taken from the oxidation tank instead of the desupersaturation tank as shown. Improvement in SO_2 removal efficiency, if any, was too small to observe.

Oxidation Efficiency - In this whole test block, the 3-inch pipe was used for discharging air into the oxidation tank and an oxidation tank level of 18 feet was maintained. Runs 820-1A, B, and C were a series to explore the air stoichiometry required to achieve near complete oxidation. Results were as follows:

Run	Air Stoichiometric Ratio, atoms O/mole SO_2 absorbed	Percent Sulfite Oxidation
820-1A	1.7	98
820-1B	1.1	92
820-1C	0	36

The oxidation efficiency was marginally acceptable at an air stoichiometric ratio of 1.1. Although the oxidation efficiency averaged 92 percent, it fluctuated widely, indicating that barely enough air was available. During the last 40 hours of Run 820-1B, air stoichiometric ratio increased to 1.3 and the oxidation efficienc was steady at 98 percent. Thus, sulfite oxidation efficiency appears to be unaffected, if not improved, by the addition of magnesium oxide.

Filter cake solids concentration at 98 percent oxidation averaged 85 percent, demonstrating that magnesium oxide addition does not adversely affect dewatering characteristics of oxidized sludge. This series of runs also demonstrated the effect of forced oxidation on solids dewatering characteristics. Filter cake solids concentration decreased from 85 percent to 63 percent as the oxidation of sulfite decreased from 98 percent to 36 percent.

An additional observation in this series of runs was that overall limestone utilization decreased from 92 to 82 percent as the air rate to the oxidation tank was reduced from 210 to 0 scfm. Presumably, the higher air rate gave better agitation of the slurry and promoted the limestone dissolution.

Section 3

FORCED OXIDATION WITH ONE SCRUBBER LOOP ON THE TCA SYSTEM

Forced oxidation with good SO_2 removal in a single scrubber loop has been demonstrated in the TCA system using limestone slurry. In this arrangement, sulfite oxidation is achieved by contacting the slurry with air in the scrubber hold tank. A compromise must be made in the scrubber liquor pH between a higher pH desired for good SO_2 removal and a lower pH desired for good oxidation. Although the optimum oxidation rate occurs at about 4.5 pH, it has been found that the oxidation rate is adequately fast up to a pH of about 6. Thus, the oxidation pH range is compatible with the limestone scrubbing pH range of 5 to 6.

Forced oxidation in a single scrubber loop is detrimental to lime slurry scrubbing because sulfite ion, a major scrubbing species in a lime based scrubbing system, is removed in the oxidation process. Thus, forced oxidation substantially reduces SO_2 removal efficiency in a single loop lime system.

The single loop configuration is of prime interest commercially because the majority of commercial installations, both operating and planned, are of this type. Modification of these installations for forced oxidation would require as a minimum a compressor (or blower) plus an air sparger in the scrubber hold tank.

Two devices for air/slurry contact have been tested on the TCA system. From late June through early October 1977 an air eductor was tested. Experience with the air eductor has been previously reported.3) Because of erosion problems and high energy consumption, the eductor has been replaced with an air sparger similar to the one used in the venturi oxidation tank. Tests with the air sparger were conducted from early December 1977 through late January 1978. All tests were conducted with flue gas containing high fly ash loadings.

SYSTEM DESCRIPTION

Two operating configurations were used in the single loop tests. With one hold tank as shown in Figure 4, effluent slurry from the scrubber is discharged to the oxidation tank where limestone is added and the slurry is recycled back to

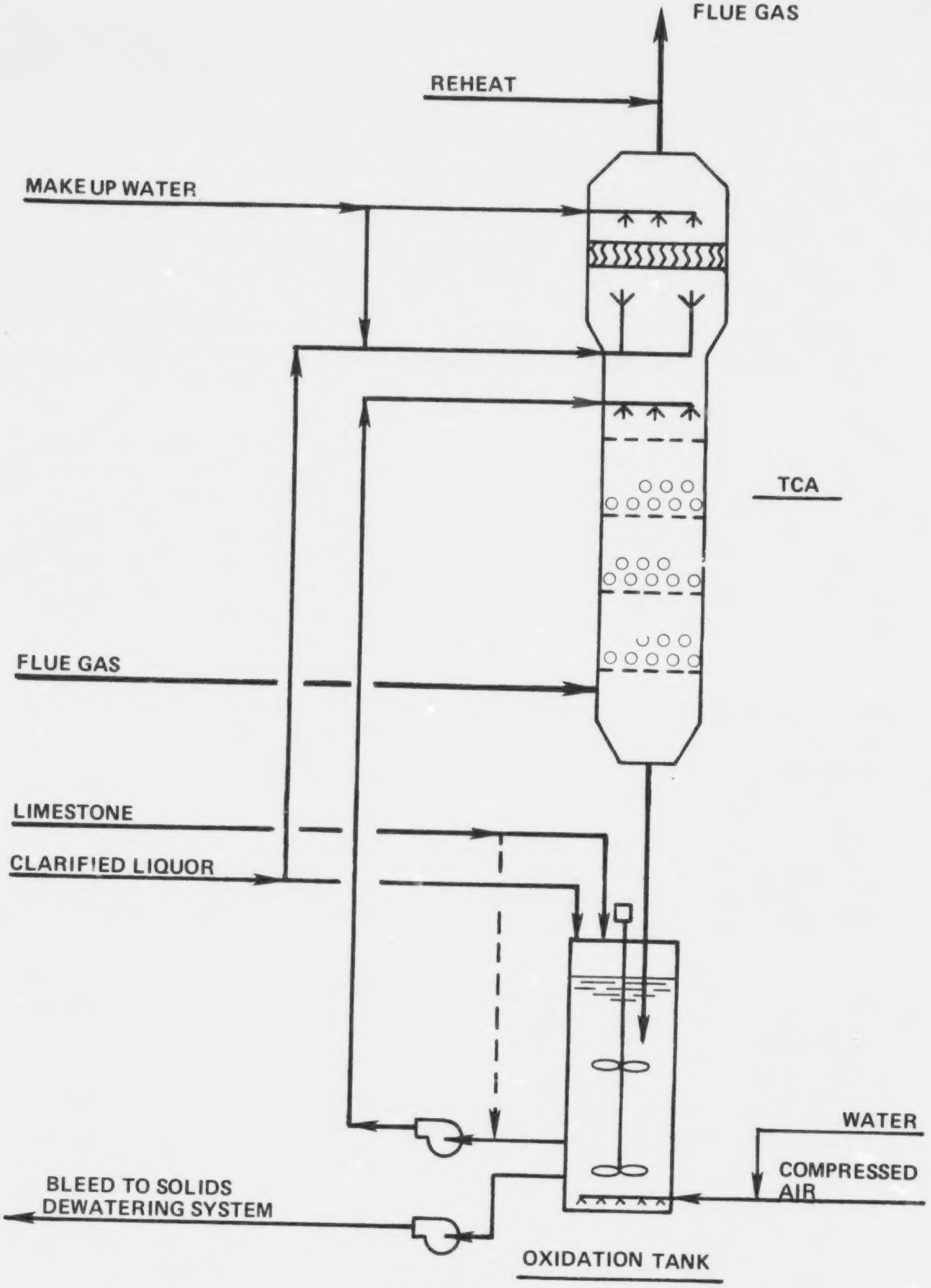

FLUE GAS
REHEAT
MAKE UP WATER
TCA
FLUE GAS
LIMESTONE
CLARIFIED LIQUOR
WATER
COMPRESSED AIR
BLEED TO SOLIDS DEWATERING SYSTEM
OXIDATION TANK

FIGURE 4. FLOW DIAGRAM FOR SINGLE LOOP FORCED OXIDATION IN THE TCA SYSTEM WITH ONE TANK

the scrubber. With two tanks in series as shown in Figure 5, effluent slurry is discharged to the oxidation tank and the slurry then passes to a second tank where limestone is added. Slurry is recycled from the second tank back to the scrubber. Although the one-tank configuration is simpler, the two-tank configuration allows the oxidation to take place at the lower pH of the scrubber effluent before limestone is added. The two-tank configuration also provides longer residence time for better limestone utilization.

The oxidation tank arrangement is shown in Figure 6. The tank is 7 ft in diameter and was operated at a 17 to 18-ft level. All tests were conducted with an air sparger ring made of straight 3-inch 316L SS pipe pieces welded into an octagon of approximately 4-ft diameter. It was located 8 inches from the bottom of the tank and had 40 1/4-inch diameter holes pointed downward. The sparger ring was fed with compressed air to which sufficient water was added to assure humidification.

A major shortcoming of this oxidation system was the agitator which was rated at only 3 Hp and rotated at 37 rpm (compared with 17 brake Hp and 56 rpm for the venturi oxidation tank). This agitator was similar in configuration to the agitator in the venturi oxidation tank with two axial flow turbines (49 inches in diameter) pumping downward. Because of the weaker agitation, runs with similar oxidation tank environment (pH, air stoichiometry, tank level, percent slurry solids, and limestone utilization) had lower oxidation efficiency in the TCA oxidation tank than in the venturi oxidation tank.

A 20 Hp variable speed agitator is on order and will be used to develop the relationship between oxidation tank agitation and air requirements.

A second shortcoming was the existing Shawnee air compressor which did not have sufficient capacity to serve the venturi and the TCA oxidation tanks simultaneously at full flue gas load. To circumvent this problem, several of the TCA runs were made at reduced flue gas flow rates. An additional air compressor has been ordered to correct this limitation.

A clarifier was used for dewatering in all runs except Run 821-2A where a clarifier followed by a centrifuge was used.

SUMMARY OF PREVIOUSLY REPORTED TEST RESULTS WITH AIR EDUCTOR

Forced oxidation test results with one scrubber loop conducted from late June through early October 1977 with limestone slurry using the air eductor have been previously reported.3) These tests were conducted at 30,000 acfm (300°F) flue gas rate which corresponds to a superficial gas velocity in the TCA of 12.5 ft/sec. The slurry recirculation rate was 1200 gpm. Each run averaged about 5 to 6 days. All runs were made with flue gas having high fly ash loadings.

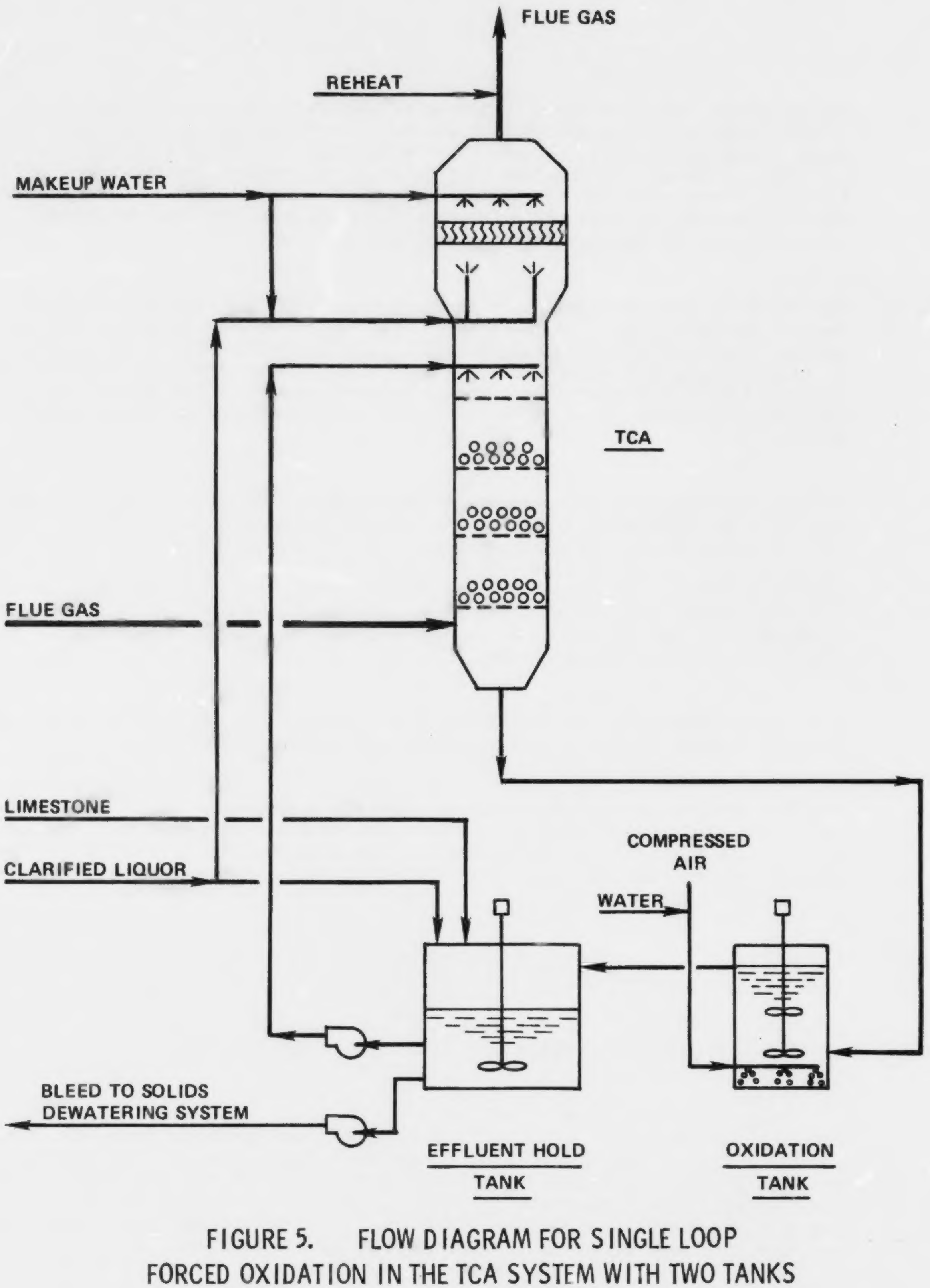
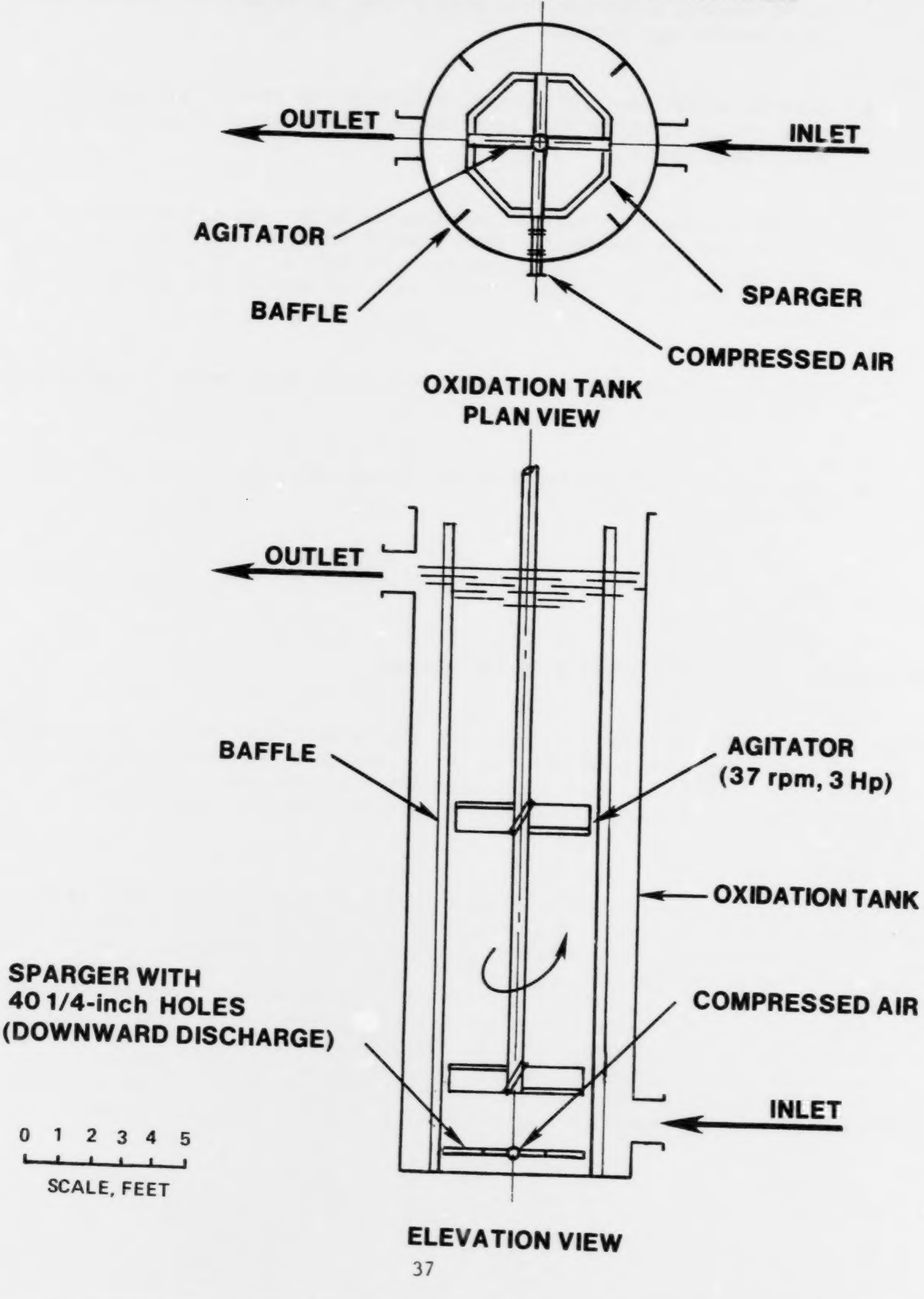

FLUE GAS
REHEAT
MAKEUP WATER
TCA
FLUE GAS
LIMESTONE
CLARIFIED LIQUOR
COMPRESSED AIR
WATER
BLEED TO SOLIDS DEWATERING SYSTEM
EFFLUENT HOLD TANK
OXIDATION TANK

FIGURE 5. FLOW DIAGRAM FOR SINGLE LOOP FORCED OXIDATION IN THE TCA SYSTEM WITH TWO TANKS

FIGURE 6

ARRANGEMENT OF THE TCA OXIDATION TANK WITH AIR SPARGER


OUTLET
INLET
AGITATOR
BAFFLE
SPARGER
COMPRESSED AIR
OXIDATION TANK
PLAN VIEW
OUTLET
BAFFLE
AGITATOR
(37 rpm, 3 Hp)
OXIDATION TANK
SPARGER WITH
40 1/4-inch HOLES
(DOWNWARD DISCHARGE)
COMPRESSED AIR
INLET
0 1 2 3 4 5
SCALE, FEET

ELEVATION VIEW

Key results from these earlier tests were as follows:

- The dewatering and handling characteristics of slurry solids oxidized to 90 percent or better in a single loop system were as good as those in a double loop system.
- Sulfite oxidation to 98 percent with good SO_2 removal was demonstrated in a single scrubber loop with two hold tanks using an air eductor for air/slurry contact.
- Conditions under which near complete oxidation was demonstrated were slurry feed to the eductor from a small downcomer hold tank at 5.15 pH, eductor discharge to the oxidation tank held at 8-ft slurry level and 5.5 pH, and an air stoichiometric ratio of about 2.5 atoms O/mole SO_2 absorbed.
- SO_2 removal was enhanced slightly by single loop forced oxidation with limestone scrubbing.
- The rubber lined eductor diffuser eroded severely in less than 1500 hours of operation.

ONE-SCRUBBER-LOOP TEST RESULTS WITH AIR SPARGER

Eight forced oxidation runs with limestone slurry were made on the TCA system in a one-scrubber-loop configuration with an air sparger. Results of these tests are reported in Table 4. Despite agitator and air compressor limitations, forced oxidation with an air sparger in a single scrubber loop was demonstrated.

Air Stoichiometry - Runs 815-2A through 818-2B were made with two hold tanks in series as shown in Figure 5. The primary effort during these runs was to identify the air stoichiometric ratio required for near complete oxidation. In the first two tests, run at the maximum achievable flue gas flow rate of 30,000 acfm, it was found that the air compressor did not have a high enough capacity to supply both the venturi/spray tower system and the TCA system. With an air rate of 210 scfm to the venturi oxidation tank only 180 scfm was available for the TCA system. Further tests were conducted at reduced flue gas flow rates (20,000 to 25,000 acfm) to allow higher air stoichiometry at the available air rate. Results of these tests conducted over an oxidation tank pH range of 5.4 to 5.7 were as follows:

Table 4

RESULTS OF FORCED OXIDATION TESTS
WITH ONE SCRUBBER LOOP ON THE TCA SYSTEM USING LIMESTONE SLURRY

Major Test Conditions	815-2A	816-2A	817-2A	818-2A	818-2B	819-2A	820-2A	821-2A
Fly ash loading	High	High	High	High	High	High	High	High
Flue gas rate, acfm @ 300°F	30,000	30,000	20,000	25,000	25,000	20,000	20,000	30,000
Slurry flow rate to TCA, gpm	1000	1000	1000	1000	1000	1000	1000	1200
Percent solids recirculated	15	15	15	15	15	15	15	15
Residence times, min: Oxidation tank	5.2	5.2	5.2	5.2	5.2	4.9	4.9	4.1
EHT	14.4	14.4	14.4	14.4	14.4	-	-	-
Oxidation tank level, ft	18	18	18	18	18	17	17	17
Airflow rate to sparger, scfm(1)	130	180	130	130	0	130	130	170
Limestone stoichiometric ratio (controlled)	1.3	1.3	1.3	1.3	1.3	1.3	-	1.2
TCA inlet pH (controlled)	-	-	-	-	-	-	5.9	-
Effective Mg^{++} concentration, ppm	-	-	-	-	-	-	-	5000
Limestone addition point	EHT	EHT	EHT	EHT	EHT	Oxid. Tk	Oxid. Tk	Oxid. Tk
Total static height of spheres, inches	20	22.5	22.5	22.5	22.5	22.5	22.5	15
Selected Results								
Onstream hours	75	48	161	131	140	164	259	182
Percent SO_2 removal	89	91	79	85	82	75	79	84
Inlet SO_2 concentration, ppm	3000	2850	3000	3000	3300	2800	2500	2500
Percent sulfite oxidation	40	54	94	67	24	94	92	95
Air stoichiometry, atoms O/mole SO_2 absorbed	1.0	1.40	1.70	1.25	0	1.90	2.0	1.65
TCA inlet pH	6.25	6.25	5.8	6.2	6.2	5.55	5.65	5.35
Oxidation tank pH	-	5.7	5.45	5.65	-	5.55	5.65	5.35
Limestone utilization, %	80	76	81	77	81	77	62	79
Gypsum saturation in TCA inlet liquor, %	110	100	100	95	110	110	115	110
Mist eliminator restriction, %(3)	0.5	-	-	-	0	-	0	1.5

Notes:

1) Air discharged downward through sparger ring with 40-¼ inch diameter holes about 8 inches from tank bottom.
2) Clarifier used for solids dewatering except for Run 821-2A where clarifier and centrifuge was used.
3) Continuous mist eliminator bottom wash with diluted clarified liquor at 0.4 gpm/ft² for Runs 815-2A & 816-2A and at 0.3gpm/ft² for Runs 817-2A through 820-2A. Intermittent bottom wash with makeup water for Run 821-2A at 1.5 gpm/ft² for 4 minutes each hour. Sequential top wash for all runs using makeup water with one of 6 nozzles on at 0.55 gpm/ft² for 3 minutes every 10 minutes.

Run	Air Stoichiometric Ratio atoms O/mole SO_2 absorbed	Percent Sulfite Oxidation
817-2A	1.7	94
816-2A	1.4	54
818-2A	1.25	67
815-2A	1.0	40
818-2B	0	24

Thus, with two hold tanks in series, an air stoichiometric ratio of about 1.7 was required to achieve greater than 90 percent oxidation. Under similar conditions in the venturi oxidation tank, higher oxidation efficiency was achieved. This better performance in the venturi oxidation tank was attributed to the superior agitation in the venturi tank.

Runs 819-2A and 821-2A were made with the oxidation tank as the only hold tank as shown in Figure 4. In these runs, the pH in the oxidation tank was higher because of the limestone addition. Because of the higher pH, a higher air stoichiometry was required. This effect can be seen by comparing Runs 817-2A (2 hold tanks) and 819-2A (1 hold tank) made at essentially the same operating conditions. Ninety-four (94) percent sulfite oxidation was achieved in both runs. An air stoichiometric ratio of 1.7 atoms O/mole SO_2 absorbed was used in the run with two hold tanks (5.4 oxidation tank pH) while an air stoichiometric of 1.9 was required in the run with one hold tank (5.65 oxidation tank pH).

SO_2 Removal Efficiency - SO_2 removal efficiency in these runs appeared to be independent of oxidation efficiency. SO_2 removal efficiency was primarily a function of flue gas flow rate and inlet SO_2 concentration, closely following previously developed correlations for the TCA system in limestone service without forced oxidation. At 3000 ppm inlet SO_2 concentration and a limestone stoichiometric ratio controlled at 1.3 moles Ca/moles SO_2 absorbed, SO_2 removal efficiency ranged from about 90 percent at 30,000 scfm to about 80 percent at 20,000 scfm.

Limestone Utilization - Limestone utilization was higher in the runs using two tanks in series than in the single tank runs. Again comparing runs 817-2A and 819-2A, limestone utilization with one tank was 77 percent while with two tanks it was 81 percent. SO_2 removal efficiency was also improved from 75 percent with one tank to 79 percent with two tanks. The improvement can be attributed to higher residence time (19.6 minutes with two tanks versus 4.9 minutes with one tank) and the approach to plug flow reaction inherent with tanks in series.

Because of the relatively poor SO_2 removal efficiency in Run 819-2A, the next run (820-2A) was made at a slightly higher pH. The oxidation tank pH was increased from 5.55 to 5.65. SO_2 removal increased only slightly from 75 percent at 2800 inlet ppm to 79 percent at 2500 inlet ppm. However, the limestone utilization decreased from 77 percent to 62 percent.

Magnesium Oxide Addition - The addition of magnesium oxide should not enhance SO_2 removal in a scrubber loop with forced oxidation. This was demonstrated in Run 821-2A. Magnesium ion in the scrubber liquor improves SO_2 removal by increasing the sulfite ion, an effective SO_2 scrubbing component. But forced oxidation converts the sulfite to sulfate which is non-reactive.

In Run 821-2A, with 5000 ppm effective magnesium ion concentration and with forced oxidation, the SO_2 removal efficiency averaged 84 percent, no higher than expected without magnesium oxide addition. In a previous run with magnesium oxide addition equivalent to Run 821-2A but without forced oxidation, SO_2 removal averaged 92 percent. Thus, the enhancement on SO_2 removal with magnesium oxide addition is not achieved in a scrubber loop with forced oxidation.

Section 4

FORCED OXIDATION OF THE VENTURI/SPRAY TOWER BLEED STREAM

Forced oxidation within the scrubber loop requires a compromise between the conditions needed for good oxidation and those needed for good SO_2 removal. Such would not be the case if it were possible to oxidize the slurry bleed stream by simple air/slurry contact. Unfortunately, tests at the IERL-RTP pilot plant2) and at the Shawnee Test Facility3) have shown that air sparging of the bleed stream increases the rate of dissolution of the residual alkali and causes the pH to rise, slowing down the oxidation rate to an impractical level. Furthermore, tests conducted with sulfuric acid addition to control the bleed stream pH have produced oxidized sludge with inferior dewatering and handling characteristics.3)

Despite the generally unfavorable results, batch oxidation tests at the Shawnee Laboratory indicated that near complete sulfite oxidation could be achieved by simple air sparging of lime or limestone slurry when magnesium ion was present in concentrations of 1,000 ppm or higher. Magnesium ion apparently has two effects: it tends to buffer the pH rise from dissolving residual alkali in the waste slurry solids; and it tends to promote dissolved sulfite availability, allowing oxidation to take place at a higher pH.

Starting in mid-May 1978, bleed stream forced oxidation with limestone slurry and added magnesium oxide was successfully demonstrated in a month long series of tests on the venturi/spray tower system. Oxidized slurry from these tests had good dewatering properties with filter cake solids concentration averaging about 85 percent. Thus, it is commercially feasible to improve the quality and reduce the volume of waste solids in installations incorporating magnesium ion in the slurry liquor by simple air/slurry contact of the bleed stream.

SYSTEM DESCRIPTION

The venturi/spray tower system was arranged as shown in Figure 7 for the bleed stream oxidation tests. Both the venturi and the spray tower slurries discharged into a single hold tank to which limestone and magnesium oxide were fed. A bleed stream was taken from the spray tower downcomer to take advantage of the low pH

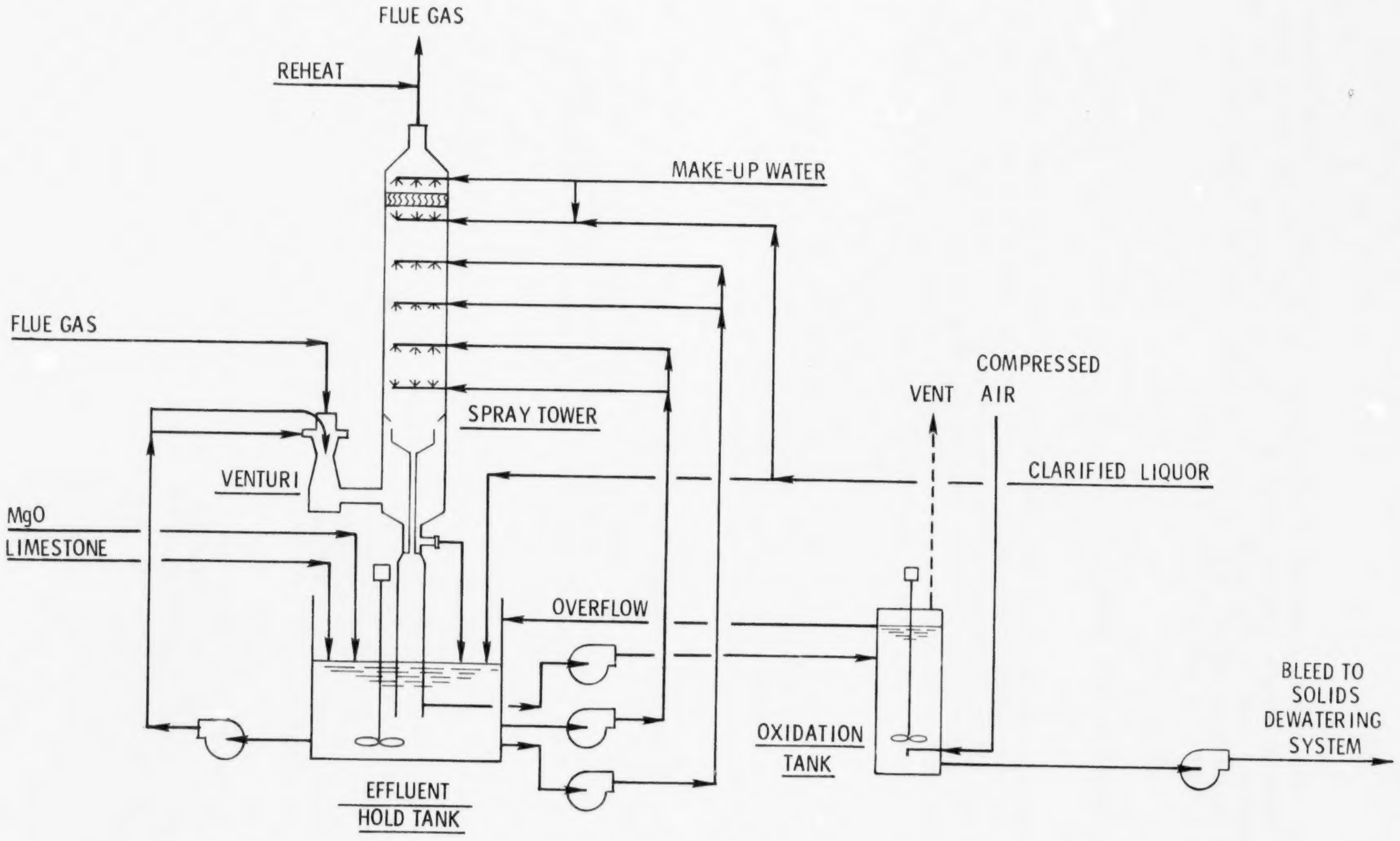

FLUE GAS
REHEAT
MAKE-UP WATER
FLUE GAS
COMPRESSED
AIR
VENT
SPRAY TOWER
VENTURI
CLARIFIED LIQUOR
MgO
LIMESTONE
OVERFLOW
BLEED TO
SOLIDS
DEWATERING
SYSTEM
OXIDATION
TANK
EFFLUENT
HOLD TANK

FIGURE 7. FLOW DIAGRAM FOR BLEED STREAM OXIDATION IN THE VENTURI/SPRAY TOWER SYSTEM

at that point. The bleed stream was discharged to the oxidation tank which was arranged as shown in Figure 3 and is described in Section 2. During these tests, the 3-inch pipe was used to discharge air into the oxidation tank. All tests were conducted at an 18-ft oxidation tank level. Bleed from the oxidation tank was dewatered by a clarifier and a filter in series.

BLEED STREAM OXIDATION TEST RESULTS

Four bleed stream oxidation runs were made on the venturi/spray tower system using limestone with added magnesium oxide. All tests were conducted with approximately 5000 ppm effective magnesium ion concentration in the slurry liquor. Test results are reported in Table 5. Percent SO_2 removal was high as expected in runs with magnesium oxide enhancement. Oxidized slurry solids in all runs had good dewatering properties, averaging about 85 percent filter cake solids concentration.

In Runs 823-1A and 824-1A, conducted at 18,000 acfm and 35,000 acfm, respectively, 97 to 98 percent sulfite oxidation was achieved at an air stoichiometry of about 1.6 atoms O/mole SO_2 abosorbed. Oxidation was consistantly high even though the oxidation tank pH averaged 6.3 in Run 823-1A and at times rose as high as 6.7.

In these runs, 30 gpm of oxidized slurry was recycled from the oxidation tank back to the scrubber hold tank. The purpose of this recycle was to reduce the pH difference between the oxidation tank and the scrubber hold tank. However, the opposite occurred. The hold tank pH was depressed, requiring excess limestone feed to maintain a pH of 5.3. The net result was a limestone utilization of less than 40 percent for these runs.

Runs 825-1A and 826-1A (at 18,000 acfm and 26,500 acfm, respectively) were conducted without this recycle. In both of these runs the pH difference between the scrubber hold tank and the oxidation tank was only about 0.1 and 0.2 with the oxidation tank pH averaging 5.65 or less. Near complete oxidation (high 90's) was easily achieved in both runs with air stoichiometric ratios of 1.6 and 2.0, respectively. Time was not available in the test block to determine minimum air stoichiometry. Control of limestone feed was poor in these runs resulting in relatively low limestone utilization (64 and 61 percent, respectively).

This short series of runs has shown that in systems containing magnesium ion, the slurry bleed stream can be readily oxidized. Furthermore, oxidation of the bleed stream does not interfere with enhancement of SO_2 removal by the magnesium ion as was experienced when oxidation was accomplished within the scrubber loop.

Table 5

RESULTS OF FORCED OXIDATION TESTS ON THE VENTURI/SPRAY TOWER BLEED STREAM USING LIMESTONE SLURRY WITH ADDED MAGNESIUM OXIDE

Major Test Conditions	823-1A	824-1A	825-1A	826-1A
Fly ash loading	High	High	High	High
Flue gas rate, acfm @ 300°F	18,000	35,000	18,000	26,500(4)
Slurry rate to venturi, gpm	600	600	600	600
Slurry rate to spray tower, gpm	1600	1600	1600	1600
Percent solids recirculated (controlled)	15	15	15	15
EHT residence time, min.	11.2	11.2	11.2	11.2
Spray tower inlet pH (controlled)	5.3	-	-	-
Scrubber limestone stoichiometric ratio (controlled)(based on solids)	-	1.9	1.4	1.4
Effective Mg^{++} concentration, ppm	5000	5000	5000	5000
Venturi pressure drop, in. H_2O	9	9	9	9
Oxidation tank level, ft	18	18	18	18
Air rate to oxidation tank, scfm(1)	110	210	110	210
Recycle flow from oxidation tank to EHT, gpm	30	30	0	0
Selected Results				
Onstream hours	205	159	229	246
Percent SO_2 removal	94	88	95	89
Inlet SO_2 concentration, ppm	2600	2600	2500	2750
Scrubber percent solids recirculated	13.3	14.1	14.7	15.2
Scrubber inlet liquor pH	5.25	5.25	5.45	5.35
Oxidation tank pH	6.30	5.90	5.65	5.45
Limestone utilization, % (based on total slurry)	36	38	64	61
Sulfite oxidation in oxidation tank, %	98	97	97	96
Sulfite oxidation in scrubber inlet slurry, %	86	49	39	29
Gypsum saturation in scrubber inlet liquor, %	120	85	105	105
Gypsum saturation in oxidation tank, %	115	90	115	115
Effective Mg^{++} concentration in scrubber inlet liquor, ppm	4990	5215	5380	4970
Oxidation tank liquor sulfite concentration, ppm	65	105	220	230
Air stoichiometry, atoms O/mole SO_2 absorbed	1.55	1.60	1.60	2.00
Filter cake solids, wt% (2)	83	85	85	84
Mist eliminator restriction, % (3)	0	0	0.5	0.1

Notes:

1) Air discharged downward through 3-inch diameter pipe with an open elbow at center of oxidation tank about 3 inches from tank bottom.
2) Clarifier and filter in series used for solids dewatering in all runs.
3) Continuous mist eliminator bottom wash with diluted clarified liquor at 0.4 gpm/ft^2 (0.3 gpm/ft^2 for Run 823-1A). Sequential top wash with makeup water with one of 6 nozzles on at 0.53 gpm/ft^2 for 4 minutes every 80 minutes.
4) Desired flow rate was 35,000 acfm but problems with the venturi lifting mechanism limited the rate to 26,500 acfm.

Additional testing will be conducted to fully characterize the limestone/MgO bleed stream oxidation system. Better operational control is required to improve the limestone utilization. Higher limestone utilization, however, is not expected to have an adverse effect on the oxidation efficiency because of the reduced amount of residual alkali and the correspondingly less possibility of pH rise in the oxidation tank.

The possibility of bleed stream oxidation on a lime/MgO system must also be investigated.

Section 5

DEWATERING CHARACTERISTICS OF THE OXIDIZED SLURRY SOLIDS

The settling and dewatering characteristics of slurry solids are routinely monitored in the Shawnee laboratory by cylinder settling tests and vacuum funnel filtration tests. Results of these monitoring tests are presented in this section. The test results are summarized in Table 6.

Results of sludge disposal studies at the Shawnee Test Facility are presented on a separate paper by the Aerospace Corporation.

Cylinder settling tests are performed in 1000 ml cylinder containing a rake which rotates at 0.16 rpm. The initial settling rate and ultimate settled solids concentration are recorded as indices of dewatering characteristics. The initial settling rate is a qualitative index of the solids settling properties only. Design rates for sizing clarifiers must take into consideration the hindered settling rate as the solids concentrate. The ultimate settled solids from the cylinder tests represent the highest achievable solids concentration in a settling pond.

Funnel filter tests are performed in a Buchner funnel with a Whatman 2 filter paper under a vacuum of 25 in. Hg. The funnel tests correlate well with the Shawnee rotary drum vacuum filter when not blinded but the funnel test cakes tend to have lower solids concentrations.

As can be seen in Table 6 the benefits of forced oxidation are clear, the dewatering characteristics of oxidized sludge are markedly better than those of unoxidized sludge. The initial settling rate is higher by a factor of 4, and both the settled and filtered solids concentrations are higher by a factor of 1.4.

Without forced oxidation, the average initial settling rate was about 0.2 cm/min. With forced oxidation the average initial settling rate was higher, ranging from 0.42 cm/min. to 1.20 cm/min.

The presence of magnesium ion tended to decrease the initial settling rate - slightl with oxidized slurry and more with unoxidized slurry. For oxidized slurry in the 2-loop mode of oxidation, the average initial settling rate was 1.0 cm/minute

Table 6

SUMMARY OF THE DEWATERING CHARACTERISTICS OF THE SHAWNEE WASTE SLURRY

Oxidation	Fly Ash Loading	Alkali	Oxidation Mode	Initial Settling Rate, cm/min		Ultimate Settling Solids, wt%		Funnel Test Cake Solids, wt%		Slurry Solids	Effective Mg^{++} Concentration, ppm
				Avg.	Range	Avg.	Range	Avg.	Range		
Yes	High	LS	1-loop	1.06	0.63-1.27	74	67-84	76	73-80	15	0
Yes	High	LS	2-loop	0.70(1) 1.20(2) 1.12(3)	0.55-0.87(1) 0.96-1.41(2) 0.81-1.47(3)	72	62-86	72	65-88	15	0
Yes	High	LS	Bleed Stream	0.42	0.29-0.61	71					
Yes	High	LS	2-loop	0.75	0.23-1.19	66	61-84	73	71-76	15	5000
Yes	High	L	2-loop	0.98	0.77-1.23	73	46-73	70	46-76	15	8000
Yes	Low	LS	2-loop	0.88	0.61-1.15	74	61-85	71	64-78	15	0
Yes	Low	L	2-loop	1.20	0.45-2.44	70	61-87	73	64-82	15	0
No	High	LS	-	0.20	0.07-0.54	54	60-81	76	64-83	15	0
No	High	LS	-	0.20	0.07-0.37	45	41-67	57	48-66	15	0
No	High	LS	-	0.05	0.01-0.11	41	30-60	57	45-64	15	5000
No	High	L	-	0.20	0.19-0.49	50	32-46	55	47-69	15	9000
No	High	L	-	0.79	0.22-1.15	42	48-66	53	51-55	15	0
No	Low	LS	-	0.17	0.05-0.47	43	31-52	52	43-63	8	2000
No	Low	L	-	0.35	0.09-0.87	40	33-54	50	41-59	15	0
							30-55	45	40-50	8	0

Note: Values for forced oxidation runs are only from data where solids oxidation was greater than or equal to 90 percent.

(1) Oxidizer pH = 4.5
(2) Oxidizer pH = 5.0
(3) Oxidizer pH = 5.5

without magnesium and 0.75 cm/min. with 8000 ppm effective magnesium ion concentration. For unoxidized limestone slurry with high fly ash loading, the average initial settling rate was reduced from 0.20 cm/min. without magnesium to 0.05 cm/min. with 9000 magnesium ion, a decrease by a factor of 4. This magnesium effect is probably the result of an increase in liquor viscosity and density due to the increased amount of total dissolved solids.

Figure 8 is a plot of percent oxidation versus settling rate for a lime system with low fly ash loading. The presence of fly ash in oxidized slurry appears to decrease the settling rate slightly. For example, oxidized lime slurry with low fly ash loading has an average initial settling rate of 1.20 cm/min., whereas oxidized lime slurry with high fly ash loading has an average initial settling rate of 0.98 cm/min.

Similar results were obtained with the ultimate settled solids and the funnel test cake solids. Without forced oxidation, the ultimate settled solids were generally in the range of 40 to 50 weight percent solids; with forced oxidation, the range was 65 to 80 weight percent. Funnel test results indicated 45 to 60 weight percent solids without forced oxidation and 65 to 85 weight percent with forced oxidation. On the rotary drum vacuum filter used in the scrubber dewatering system, cake solids concentration was always above 80 percent with oxidized slurry while averaging 50 to 60 percent with unoxidized slurry.

The unoxidized solids tended to be thioxotropic, like quicksand, while the oxidized solids were more like moist soil.

The presence of magnesium ion did not affect the ultimate settled solids or the funnel test cake solids results. This result was also seen in the test facility's operating data.

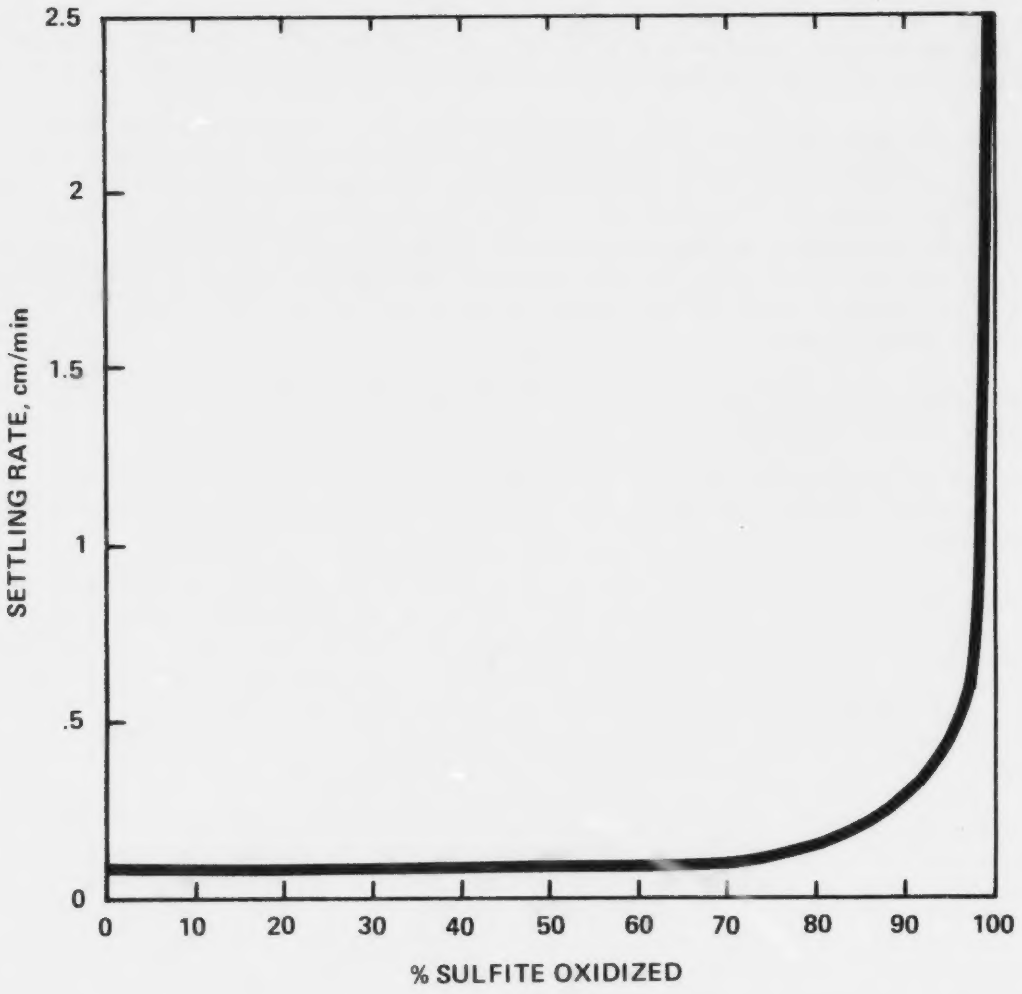
The product solids at Shawnee are a mixture of unoxidized calcium sulfite hemihydrate ($CaSO_3 \cdot 1/2H_2O$), the oxidation product calcium sulfate dihydrate ($CaSO_4 \cdot 2H_2O$), fly ash, unreacted alkali, and other inert materials. In slurry that is 10 percent oxidized, the main component is calcium sulfite in platelets and rosettes of only a few microns in diameter. The solids in slurry 95 percent oxidized are mainly calcium sulfate crystals having a bulky rectangular shape and ranging in size from 20 to 100 microns.

It is currently thought that in the case of unoxidized slurry, the calcium sulfite fines are the limiting factor in the inital settling rate. But in the case of oxidized slurry, the fly ash may be the limiting factor. Fly ash is extremely fine when compared with the calcium sulfate particles and hence settles at a slower rate. At the Shawnee Test Facility, the limit for oxidized slurry with high fly ash loading appeared to be 1.5 cm/min. For oxidized slurry with low fly ash loading, the limit was 2.4 cm/min. Since, at Shawnee, solids from flue gas with low fly ash loading contain up to 1 weight percent fly ash, the limits of the calcium sulfate settling rate may be even higher.

The residence time in the oxidation tank may also affect the size of the gypsum crystal; the longer the residence time the larger the crystal. This effect has yet to be thoroughly explored at Shawnee.

FIGURE 8

EFFECT OF OXIDATION ON INITIAL SETTLING RATE



Section 6

FUTURE TESTING

Testing with forced oxidation will be continued with emphasis on:

- More fully developing forced oxidation within a single scrubber loop
- Exploring the conditions under which bleed stream oxidation is applicable
- Determining compatibility of forced oxidation with chemical additives such as adipic acid

Based on the encouraging results at the IERL-RTP pilot plant, an extensive program to develop adipic acid as an additive for enhancing SO_2 removal efficiency has recently been initiated. Adipic acid acts as a buffer to limit the drop in pH, thereby improving the liquid-phase mass transfer. The advantages of adipic acid are listed below:

- Lower cost compared to MgO based on the quantity needed. For example, for a similar degree of SO_2 removal enhancement in limestone scrubbing:

  With MgO: 6,000 ppm Mg++ at $0.17/lb MgO requires $14/1,000 gal of discharged liquor

  With adipic acid: 1,000 ppm adipic acid at $0.42/lb acid requires only $3.50/1,000 gal of discharged liquor

- Optimum adipic acid concentration for effective improvement in SO_2 removal is only 5-10 m-moles/liter (700 - 1,500 ppm)
- Adipic acid improves SO_2 removal and may also improve limestone utilization, whereas MgO may reduce the limestone dissolution rate
- In the limestone scrubbing system, SO_2 removal efficiency is no longer limited by the limestone dissolution rate when adipic acid is present in sufficient quantity

- Forced oxidation does not affect the effectiveness of adipic acid. Forced oxidation reduces the effectiveness of the magnesium ion by converting the scrubbing $SO_3^=$ into non-scrubbing $SO_4^=$ species

- Unlike MgO addition, where two chloride ions tie up a magnesium ion to form neutral $MgCl_2$, adipic acid is not affected by chloride

- Adipic acid is nontoxic (used as a food additive)

- Both IERL-RTP pilot plant and preliminary Shawnee results show that the solids quality (filterability, settling rate) is not affected by adipic acid

Tests with adipic acid as an additive are scheduled with and without forced oxidation and with both lime and limestone slurries.

Tests are also planned to investigate the effect of limestone type and grind on SO_2 removal and limestone utilization. Initial screening tests have already been scheduled for the IERL-RTP pilot plant.

A 3-month test block is planned for the Shawnee spray tower with various internal configurations (different number of headers, number of nozzles, type of nozzles, nozzle pressure drop, etc.). The primary objective of the testing will be to provide a better basis for designing full-scale spray towers.

Long-term (over one month) lime and limestone tests are also planned, which will combine the most promising operating conditions, including forced oxidation and organic acid addition, to demonstrate system reliability and conformance to the existing Federal emission standard.

Other conc[illegible]ent future activities include:

- Transfer of Shawnee-developed technology to full-scale plants, including, if necessary, simulation of commercial plant operation at Shawnee

- Continued development and updating of the Economic Study Computer Program in conjunction with TVA

- A study of the overall power plant water management as it relates to FGD plant operation

Section 7

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SELECTED TOPICS FROM
SHAWNEE TEST FACILITY OPERATION

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INTRODUCTION

The Shawnee Test Facility is an EPA funded wet lime/limestone scrubbing test facility that has operated since 1972. The facility currently consists of two 10 MW equivalent scrubbing systems each treating approximately 7 percent of the flue gas produced by a medium-to-high sulfur coal-fired 150 MW boiler. One system is a venturi followed by a spray tower (V/ST) the other is a Turbulent Contact Absorber (TCA).

The information presented here reflects selected topics from the operating experience obtained between June 1977 and May 1978 at this facility. These include:

a) Scrubber operation and maintenance

b) Dewatering systems

c) Forced oxidation systems

d) Automatic control of limestone addition

e) Operational development

SCRUBBER OPERATION AND MAINTENANCE

SCRUBBER OPERATION

Over the past year the V/ST and the TCA have maintained a high operational availability that is summarized in Table 1. The V/ST system operated 7040 hours during the year or 80 percent of the time. The TCA system operated 7272 hours or 83 percent of the time. The systems downtimes were attributed to four categories:

a) Boiler outages

b) Weather affecting scrubber operations

c) Scheduled scrubber system inspections and modifications

d) Scrubber equipment failures

The unit 10 boiler operated 91 percent of the time and consequently caused 9 percent downtime for each of the two systems for the year. Of the over 800 hours of boiler downtime approximately 350 hours were the result of a scheduled major overhaul. The hours of downtime differ for the two systems because the V/ST and TCA had to start-up separately due to operating personnel limitations.

TABLE 1
SCRUBBER OPERATION

	VENTURI/SPRAY TOWER		TCA	
OPERATING TIME, Hours		7040 (80%)		7272 (83%)
DOWN TIME, Hours				
BOILER & DOE	832 (9%)	1720 (20%)	810 (9%)	1488 (17%)
WEATHER	220 (3%)		171 (2%)	
INSPECTION & MODIFICATION	421 (5%)		323 (4%)	
EQUIPMENT PROBLEMS	247 (3%)		184 (2%)	
TOTAL PERIOD*, Hours		8760 (100%)		8760 (100%)

*JUNE 1977 THROUGH MAY 1978

The second major cause of downtime was attributed to routine scrubber inspections and system modifications inherently associated with the goals of this test facility. Typically a test would last 5 to 6 days with occasional runs lasting as long as 9 days to insure sufficient steady state data. The time spent between test runs varied from a few hours for a scrubber inspection to 3 or 4 days for a system modification. The factors caused downtimes for the year of 5 percent for the V/ST and 4 percent for the TCA.

The remaining two causes of outages were scrubber equipment failures and weather related problems. Equipment failures and subsequent maintenance will be discussed in depth in the next section. The weather related problems were the result of the abnormally severe winter experienced in Kentucky last January. Alkali addition streams were the most troublesome in that the one-inch utility hoses used as alkali addition lines froze repeatedly. Maintenance was seriously hampered by the cold. The downtime for the year caused by the weather was 3 percent for the V/ST and 2 percent for the TCA.

Curtailed power generation because of scrubber inavailability is of serious concern to utility companies. For this paper, scrubber availability is introduced and is defined as the percent of time that a commercial scrubber operates while the accompanying boiler is operating. Scrubber equipment failures and weather created problems were used in calculating the availability term at Shawnee. Modifications and inspections due to test requirements were not included because they were inconsistent with the concept of a commercial installation.

For the V/ST and TCA, the average availability for the last year was 94 and 95 percent, respectively. On a monthly basis the availability ranged from 82 to 100 percent for the V/ST and 81 to 100 percent for the TCA.

MAINTENANCE

At the Shawnee Facility, the high availability has been strongly influenced by an effective maintenance program and an adequate spare parts inventory. These factors are particularly important when considering pump maintenance; the facility does not have stand-by spare pump capacity as is common practice in industrial systems. The total hours and frequency associated with equipment maintenance that resulted in system shut-downs are outlined in Tables 2 and 3 for the V/ST and TCA, respectively.

In the V/ST system the primary cause of downtime was scrubber related problems that consisted, in part, of a corroding air sparge pipe, pipe plugging and leaks, nozzle plugging, and solids build-up in the outlet ductwork. The corroding sparge pipe occurred only once and was caused by mistakenly using 304 stainless steel as part of the material of construction.

TABLE 2

V/ST EQUIPMENT MAINTENANCE CAUSING DOWN TIME

JUNE 1977 TO MAY 1978

ITEM	FREQUENCY OF EVENT	TOTAL HOURS
SCRUBBER INTERNALS AND PIPING	12	114
I. D. FAN	4	78
INSTRUMENTATION	1	39
PUMPS	2	12
ALKALI FEED	1	3
OTHER	1	1
TOTAL	21	247

TABLE 3
TCA EQUIPMENT MAINTENANCE CAUSING DOWN TIME
JUNE 1977 TO MAY 1978

ITEM	FREQUENCY OF EVENT	TOTAL HOURS
SCRUBBER INTERNALS AND PIPING	6	108
AGITATOR	1	39
PUMPS	4	17
ALKALI FEED	2	8
INSTRUMENTATION	1	7
OTHER	2	5
TOTAL	16	184

The longest downtime requirement for the TCA was the installation of the Penberthy eductor. The agitator shaft replacement in the main hold tank was second, followed by routine maintenance of the main slurry pump which consisted of repacking Allen-Sherman-Hoff Centri-Seals. Other operating problems included plugging of the alkali feed lines, bearing failures of the I.D. fan and general solids build-up in the outlet duct.

The personnel available for maintenance requirements and system modifications are listed in Table 4. To avoid possible conflicts of priorities between the powerhouse and scrubbing facility and to allow the crafts people to develop expertise with specific equipment, the maintenance personnel are assigned exclusively to the test facility.

DEWATERING SYSTEMS

The primary dewatering of the purged slurry in both scrubbers is achieved by clarifiers. Further dewatering of the clarifier sludges is accomplished in the V/ST system by a filter and in the TCA system by a centrifuge. The information discussed below represents actual data that might be used as a guide in understanding the operating problems and costs associated a with centrifuge or filter.

CENTRIFUGE

A continuous centrifuge is one process used to dewater scrubber waste sludge and to recover the dissolved scrubbing additives. The normal operating conditions usually consist of a feed stream flow of 15 gpm at 30 to 40 wt. percent solids, a centrate of 0.1 to 3.0 wt. percent solids, and a cake of 55 to 65 wt. percent solids, for unoxidized slurry. Approximately 30 percent of the total solids is fly ash; the remaining solids are predominantly calcium sulfate and sulfite.

The machine is a Bird 18" x 28" solid bowl continuous centrifuge which operates at 2050 rpm. The material of construction is 316L stainless steel with stellite hardfacing on the feed ports, conveyor tips and solids discharge ports. The bowl head plows and case plows are replaceable. The pool depth is set at 1-1/2 inches. No cake washing is performed in this machine.

The centrifuge was inspected in June 1978, after 6460 hours of operation since the previous factory servicing. The inspection was prompted by the gradual and continued increase of centrate suspended solids to a level of approximately 3 wt. percent. The machine was judged to be in generally fair condition but certain components were in need of factory repair. Serious wear was observed at the conveyor tips on the discharge end and at the junction of the cylinder and the 10° section of conveyor. Wear was also present at the casing head plows and solids discharge head near the discharge ports. The bowl and effluent head were in good condition.

TABLE 4
SHAWNEE FACILITY
MAINTENANCE PERSONNEL

I	MAINTENANCE FOREMAN	1
	BOILERMAKER – WELDER	2
	CARPENTER	1
	ELECTRICIAN	1
	PIPEFITTER – WELDER	2
	TEAMSTER	1
	LABORERS	2
	HEAVY EQUIPMENT OPERATOR	1
	PAINTER	1
	MACHINIST	1
	INSULATOR	1
II	INSTRUMENT FOREMAN	1
	SENIOR INSTRUMENT MECHANIC	2
	JOURNEYMAN INSTRUMENT MECHANIC	2
III	TOTAL	19

TABLE 5
CENTRIFUGE MAINTENANCE & POWER REQUIREMENT
JUNE 1977 TO MAY 1978

I MAINTENANCE

EVENT	FREQUENCY OF OCCURRENCE	ESTIMATED TOTAL ONSITE LABOR (MAN-HOURS)	ESTIMATED TOTAL MATERIAL COST ($)
FEED PIPE REPAIR	1	16	20

II POWER REQUIREMENT – 30 HORSEPOWER

The current plans call for the centrifuge to be shipped to the factory for an overhaul. The following items will be accomplished:

a) Inspect and service the gear and bearing unit

b) Rebuild all worn conveyor surfaces and add hardfacing on the tips

c) Rebuild and add hardfacing to the discharge ports

d) Replace all seals and bushings in the effluent and discharge head

e) Replace case plows and discharge plows as necessary

In an attempt to improve performance and machine life, tungsten carbide hardfacing will be applied to the conveyor tips instead of the previously used stellite. The estimated cost for the complete factory service, including the hardfacing, is $17,000.

The machine has been a minimum maintenance item; as shown in Table 5 the only maintenance of the past year has been the replacement of the feed pipe in the centrifuge. The power requirement for the machine is 30 horsepower.

FILTER

An Ametek 3' x 6' vacuum drum filter without cake wash is operated at the facility for waste sludge dewatering and dissolved scrubbing additive recovery. The feed to the filter is usually 15 gpm of 30 to 40 wt. percent solids.

The filtrate generally contains less than 0.02 wt. percent solids. The filter cake varies from 55 to 85 wt. percent solids depending mainly on whether the sludge is unoxidized or oxidized.

The filter, with the exception of the filter cloth, has been a moderate maintenance item. Table 6 is a breakdown of maintenance categories, frequency, approximate total manhours required, and approximate replacement material costs. Also, included in Table 6 is the power requirement for the filter and vacuum pump.

Contrary to experiences at other scrubbing facilities, filter cloth replacement as noted in Table 7 has been a serious problem at Shawnee. The causes of cloth blinding and fraying are not satisfactorily understood as yet. However, in the last few months operating experience indicates a relationship between cloth

TABLE 6
FILTER MAINTENANCE & POWER REQUIREMENT
JUNE 1977 TO MAY 1978

I MAINTENANCE

EVENT	FREQUENCY OF OCCURRENCE	ESTIMATED TOTAL ONSITE LABOR (MAN-HOURS)	ESTIMATED TOTAL MATERIAL COST ($)
SPEED CONTROL REPAIR	2	16	1000
CAKE DISCHARGE AIR REPAIR	2	16	0
CLOTH REPLACEMENT	15	90	1500

II POWER REQUIREMENT – 20 HORSEPOWER

TABLE 7
FILTER CLOTH SERVICE

	CLOTH TYPE*	DATE INSTALLED	HOURS IN SERVICE	COMMENT
1	TFI	6 - 6 - 77	292	BLINDED
2	LAMPORTS	7 - 7	127	HOLE IN CLOTH
3	LAMPORTS	7 - 13	142	BLINDED
4	LAMPORTS	7 - 19	302	BLINDED
5	LAMPORTS	8 - 3	249	BLINDED
6	LAMPORTS	8 - 15	540	BLINDED
7	TFI	9 - 19	187	BLINDED
8	TFI	9 - 23	501	BLINDED
9	LAMPORTS	10 - 18	187	BLINDED
10	AMETEK	10 - 26	2096	HOLE IN CLOTH
11	AMETEK	2 - 6 - 78	188	BLINDED
12	LAMPORTS	2 - 14	190	HOLE IN CLOTH
13	LAMPORTS	2 - 22	290	BLINDED
14	AMETEK	3 - 21	535	BLINDED
15	AMETEK	4 - 19	1197	HOLE IN CLOTH

*AMETEK – AMETEK OLEFIN (STE – F9D8 – HJO)

LAMPORTS – LAMPORTS POLYPROPYLENE (7512 – SHS)

TFI – TECHNICAL FABRICATORS INCORPORATE, POLYPROPYLENE (9162) life and the technique by which the cloth is fitted to the drum. A carefully controlled amount of looseness in the fit between the dividers appears to be desireable for cake discharge and non-blinding. The looseness evidently allows the cloth to "snap" the cake off when the air puff of the cake discharge cycle is applied to the given filter cloth section. Two additional observations have been that oxidized sludge exhibits less tendency towards cloth blinding and Ametek olefin appears to provide the most satisfactory service of these cloths tested. The reason for the better service life of Ametek is suspected to be attributable to the looseness of weave that the Ametek has in comparison with Lamports and TFI.

As noted, some limited progress has been achieved towards understanding and reducing cloth blinding. Future operations at Shawnee will include continued efforts to improve filter cake performance.

OXIDATION SYSTEM DISCUSSION

The desirability of an oxidized calcium base sludge has for a number of years been recognized when considering ease of sludge filtration, handling and disposal. Forced oxidation has been determined to be one method in achieving those ends. In forced oxidation, air is introduced and dispersed into the slurry as required to oxidize the calcium sulfite to calcium sulfate.

At Shawnee, forced oxidation has been investigated in a series of tests using a Penberthy eductor and an air sparger. The intended goal of these tests was to investigate the operating parameters necessary to achieve "near complete" oxidation, i.e., greater than 90 percent total sulfur as sulfate.

One important operating experience resulting from these tests was a tight water balance with no observable deterioration of mechanical components or scrubbing chemistry.

PENBERTHY EDUCTOR

The Penberthy eductor is a device similar in concept to a laboratory aspirator. A high velocity slurry passes through a constricted nozzle, into an eductor chamber and then through a moderately restricted jet throat. In the chamber the slurry induces a vacuum that draws ambient air into the chamber via an entry pipe located at right angles to the slurry flow path. The air is then entrapped and mixed in the slurry as the fluid leaves the eductor chamber and passes through the jet throat.

At Shawnee, a Penberthy Model ELL-10 eductor was tested. The materials of contruction were stellite for the nozzle and neoprene-lined carbon steel for the eductor chamber and exit jet throat. The system was operated at 1600 gpm.

Test results indicated that near complete oxidation could be achieved, but at the same time serious erosion problems developed. The neoprene lining in the jet throat was observed to be "chipped off" after only 620 hours of operation. After approximately 1800 hours of operation, bare carbon steel was exposed, and after an epoxy patch failed after 2055 hours of operation, the tests were terminated.

Currently, no plans exist to resume testing of eductors primarily because no advantages in oxidation capabilities were seen in comparison with sparge air systems. Secondary reasons included the materials erosion problem and the unfavorable operating and capital costs in comparison to sparge air systems.

AIR SPARGER

The sparge air system used in sludge oxidation is a simple concept that involves bubbling air into the bottom of a slurry tank in conjunction with simultaneous slurry agitation. At Shawnee the oxidation tanks are purposely tall and thin (7 to 8 feet diameter x 20 feet tall). This shape enables a long air/slurry contact time.

In the TCA an octagonal sparge ring with 40 holes is currently installed at the bottom of the effluent hold tank. The holes are 1/4 inch diameter and located on the underneath side of the ring; no sparge ring plugging had occurred to date. The ring is constructed of 316L stainless steel and is materially in good condition except for minor erosion at the 1/4 inch air holes. Currently a conventional tank agitator (37 rpm, 3 hp) is in service. Future tests will use a variable, high speed agitator. Near complete oxidation has been achieved with the current configuration.

Testing was done on the V/ST system with an octagonal sparge ring similar in design to that of the TCA. After 2400 hours of operation the ring was removed from service and replaced with a 3-inch diameter open-ended sparge pipe. The air and slurry are mixed with an axial flow agitator (56 rpm, 20 hp). Near complete oxidation has been obtainable with both the sparge ring and sparge pipe in conjunction with the agitator.

Both V/ST and TCA sparge systems operate from the same Worthington oil-free air compressor under the following conditions: 50 psig, 270 F, and flow rates normally at 210 scfm per scrubber. The compressor loading/unloading cycle is normally 4 seconds/10 seconds for 210 scfm. The oil-free compressor was chosen to minimize the possibility of slurry contamination by oil which contains oxidation inhibitors.

Maintenance on the sparge systems has been mainly confined to the compressor. The suction valve has been serviced and the compressor cooling water jacket has been flushed periodically.

Further testing of forced oxidation with the sparge ring and sparge pipe configurations is planned.

AUTOMATIC LIMESTONE ADDITION CONTROL

An automatic limestone feed control system was installed and initially tested on the TCA in April-June 1977. The control logic was based on a material balance concept of maintaining a desired stoichiometric limestone feed in relation to the amount of SO_2 absorbed in the scrubber. The results of the tests indicated that limestone addition was satisfactorily controlled during normal fluctuations of SO_2 inlet mass flow rates. As will be discussed, one observed weakness of the control system was that the stoichiometric ratio was independent of the SO_2 inlet mass flow rates and therefore the control system could not effectively compensate for unusually large fluctuations in SO_2.

The basic control scheme is represented by the following equation and by Figure 1.

$$L = G \times (S - K) \times R$$

where: L = limestone slurry addition rate, gpm

G = flue gas flow rate, acfm

S = inlet SO_2 concentration, ppm

K = a manually adjustable constant related to desired outlet SO_2 concentration, ppm

R = a manually adjustable constant proportional to the desired stoichiometric ratio

= (unit conversion factor) x (stoichiometric ratio)

= 2.34×10^{-8} x (stoichiometric ratio)

The factor G(S-K) represents the amount of SO_2 absorbed per unit time. Thus, at a set value of R which is proportional to the desired stoichiometric ratio, the limestone addition rate is automatically adjusted to maintain the desired stoichiometry.

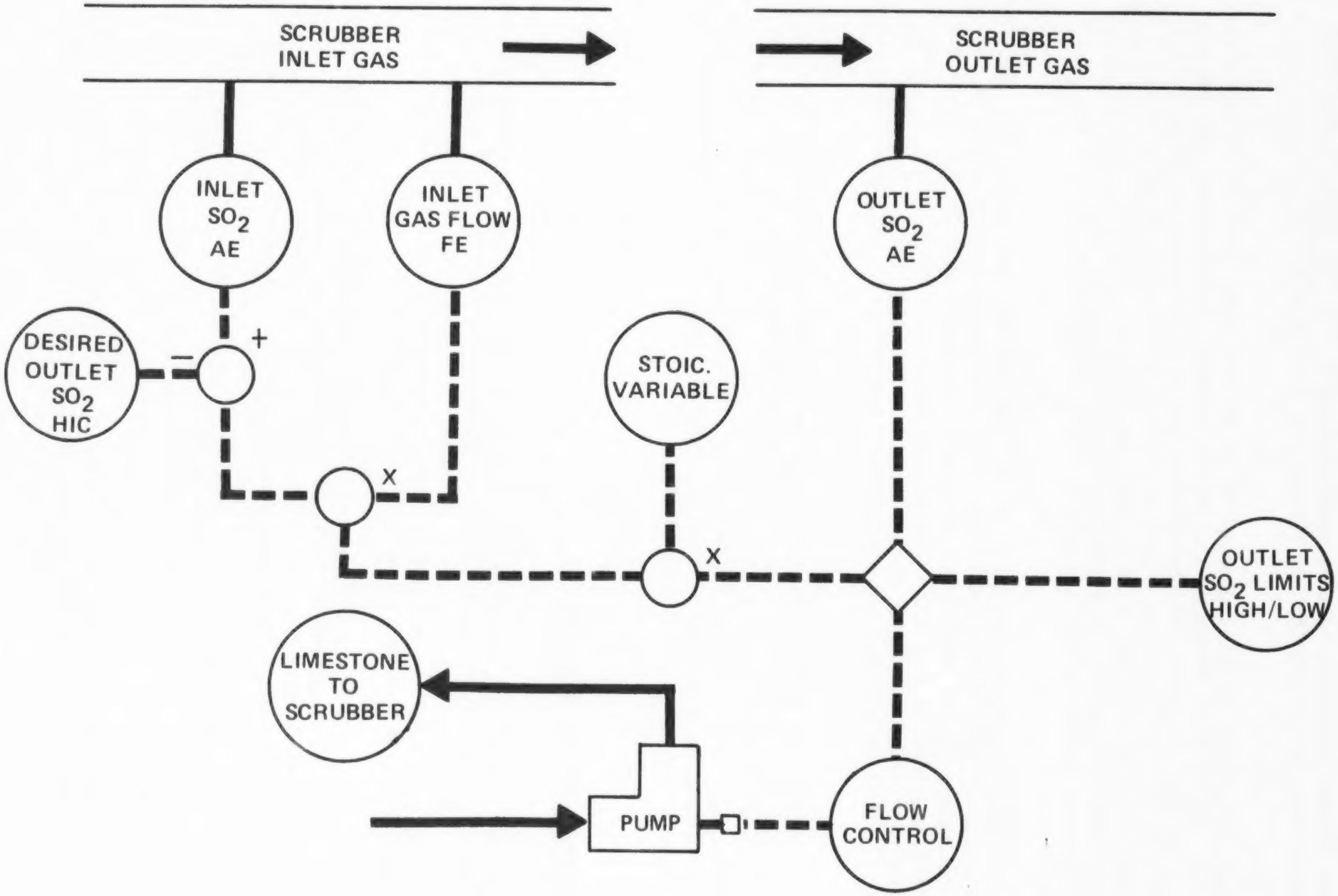
In practice, the inlet SO_2 concentration, S, and the flue gas flow rate, G, may vary within a wide range depending on the sulfur content in the coal and the boiler load. Therefore, the control scheme also includes overrides which are activated when the following situations arise:

a) If the measured outlet SO_2 concentration exceeds a set maximum, the limestone addition rate will be stepped up to a preset maximum.

b) If the measured outlet SO_2 concentration drops below a set minimum, the limestone addition rate will be maintained at a preset minimum.

FIGURE 1

ALKALI FEED CONTROL SCHEME


SCRUBBER INLET GAS
SCRUBBER OUTLET GAS
INLET SO_2 AE
INLET GAS FLOW FE
OUTLET SO_2 AE
DESIRED OUTLET SO_2 HIC
+
−
X
STOIC. VARIABLE
X
OUTLET SO_2 LIMITS HIGH/LOW
LIMESTONE TO SCRUBBER
PUMP
FLOW CONTROL

The former provision insures compliance with the SO_2 emission standard, while the latter reduces the posibility of pipe plugging.

Operating under typical conditions of 30,000 acfm, 1200 gpm liquor rate, and 3 beds each with 5 inches static height nitrile foam spheres, the desired limestone stoichiometric ratio for the TCA was set at 1.35 moles Ca added/mole SO_2 absorbed ($R = 2.34 \times 10^{-8} \times 1.35 = 3.16 \times 10^{-8}$). Actual stoichiometric ratio varied between 1.18 and 1.55, with an average of 1.37. Actual limestone addition rate was generally within 10 percent of the rate calculated from the control equation presented above. A desired outlet SO_2 concentration, K, was set at 430 ppm.

Initially, overrides were set at 500 ppm and 200 ppm outlet SO_2 concentration. Because of wide variation in the inlet SO_2 concentration (2100 to 3400 ppm), the outlet SO_2 concentration frequently exceeded the 500 ppm upper override limit. This resulted in the actuation of the override control limit and interferred with the testing of the major control logic. Subsequently, the upper override limit was raised to 800 ppm and the proportional control functioned more smoothly.

In theory, the higher the inlet SO_2 concentration, the higher the required percent SO_2 removal in order to meet the SO_2 emission standard. The higher required percent SO_2 removal would, in turn, call for higher limestone stoichiometric ratio. In the present setup, the constant R can only be reset manually. Therefore, as a second generation control logic, R as a function of the inlet SO_2 concentration would be desirable.

OPERATIONAL DEVELOPMENT

In the 6 years of operation the intent of the test facility has been to accelerate the development and application of lime/limestone scrubbing technology. In addition to better understanding the chemistry of scrubbing, improving the sludge disposal properties and enhancing SO_2 removal, countless developments and improvements have been made in the operation of the facility. A few are listed here:

a) Mist eliminator plugging at Shawnee is extremely infrequent. Wash patterns and wash sequences have been refined to such a degree that an excess of 7700 hours of operation was obtained recently between cleanings on the TCA mist eliminator. The reason for the cleaning was that 12% plugging was present and a new series of tests using adipic acid were to begin. The V/ST mist eliminator was cleaned only once during the last 8000 hours of operation due to problems with the wash system that caused 15% plugging. Considering the wide range of operating conditions for the systems, these are outstanding records.

b) The variable speed Allen-Sherman-Hoff rubber lined pumps have demonstrated their effectiveness as slurry pumps under a wide range of operating conditions.

c) The process pH is measured using Uniloc Model 321 submersible electrode assemblies. Originally, Uniloc Model 320 flow-through meters were installed. But because of line plugging problems and frequent sensing electrode breakage this type of sensor was abandoned. Current service requirements for the submersible assemblies consist of periodic cleaning and buffering of the electrodes, generally every 2 or 3 days to insure accuracy. Also to minimize the service requirement, the instrument electrodes are placed in water when the scrubbers are not operating. Years of operation have shown that to insure the accuracy of the process pH meters, a laboratory measured pH should be taken once every four hours for comparison purposes. This procedure enables a normal operation to within ± 0.2 pH units of any desired set point.

d) A Dupont Model 400 UV split-beam photometer is used to measure SO_2 concentrations In the last few years the instrument has been accurate and reasonably trouble-free. Maintenance requirements are limited to cleaning the sample cell and sample lines approximately once every 1 or 2 months and cleaning the particulate filter usually once every 3 to 4 weeks. Ultraviolet lamp failure has been the only component problem and has been caused by uncontrollable and momentary power fluctuations due to the switching of station power. The effective particulate filter for the instrument at Shawnee is a cylindrical chamber constructed of a fine mesh screen. The screen cylinder is surrounded by a solid protective cylinder. The gas sample lines have operated leak free; the lines are 316L stainless steel tubing with heat tracing.

e) Oxygen in the inlet flue gas is measured with a Teledyne Model 9500 which uses a micro-fuel cell. Operating performance has not been acceptable. A frequent problem has been the rapid deactivation of the special micro-fuel cells. Service life has varied from one day to approximately 1-1/2 months. In some cases, the cells have arrived from the factory in a deactivated condition. The causes of the cell deactivation might be due to exposure to a CO_2-free environment or factory defective cells. The problem with cell life continues and is being studied.

f) The Foxboro 2800 series and 1800 series magnetic flow meters have shown no serious problems. Periodic scale cleaning is required to improve accuracy and sensitivity but the meters are considered reliable, acceptably accurate and easily serviced.

g) Both Dynatrol Model CL-10HY U-tube density meters and Ohmart radioactive density meters are used at Shawnee. Both meters provide acceptably accurate and dependable service. From an operations point of view, the U-tube meters did have some initial problems with line plugging. The cause was attributed to operator error in setting too low a flow through in instrument. The problem has since been resolved.

In the future, operational areas of interest will include but not be limited to the continued effort to understand and control scale growth, improve dewatering equipment operation, reduce the frequency of routine maintenance, and optimize pump seal water requirements.

STATUS REPORT OF
SHAWNEE COCURRENT AND DOWA SCRUBBER PROJECTS
AND
WIDOWS CREEK FORCED OXIDATION

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Electric Power Research Institute
Palo Alto, California

Prepared for Presentation at
Industry Briefing Conference
Results of EPA Lime/Limestone Wet Scrubbing Test Programs
Sponsored by the U.S. Environmental Protection Agency
Royal Villa Motel in Raleigh, North Carolina
August 29, 1978

Introduction

In 1970, the Tennessee Valley Authority and the Environmental Protection Agency began a joint program at TVA's Shawnee Steam Plant to evaluate processes which would remove sulfur dioxide and particulates from the gaseous emissions of a coal-fired power plant. The major areas of concern investigated are: the cost of removal, the reliability of the process, the availability of materials needed and waste disposal of the byproducts from the removal processes.

Three 10-MW scrubbers were constructed for testing by TVA. Each has the capability of pulling flue gas from the No. 10 boiler either before or after the electrostatic precipitator. Simultaneous testing has continued on various lime and limestone removal processes in an attempt to lower capital and operating costs to make each process more reliable, to optimize the processes, and to stabilize or utilize the waste products from the processes.

Two advanced systems are presently being prepared for testing at a 10-MW size. The first is a cocurrent scrubber which has the potential advantage of a smaller scrubber vessel, thus lower capital cost over a conventional scrubber. Testing of the cocurrent scrubber has been completed on an 1-MW pilot plant at Colbert Steam Plant and the 10-MW prototype is in initial testing phase. The other process is the DOWA process. This Japanese process is being marketed in the U.S. by Universal Oil Products (UOP). This process can produce a stable, storable product (gypsum) and, where demanded, a sellable product.

Forced Oxidation has been demonstrated at Shawnee Steam Plant on a 10-MW facility. TVA plans to demonstrate forced oxidation on a full-scale facility at Widows Creek Steam Plant. Testing will include technical feasibility and environmental acceptability of forced oxidation as a method of sludge disposal.

Cocurrent 1-MW Pilot Plant Results

EPRI funded TVA to evaluate the cocurrent scrubber concept at the Colbert 1-MW pilot plant and to provide design data for a 10-MW prototype to be constructed and operated at the Shawnee Scrubber Test Facility.

The emphasis of the evaluation was to study (1) the gas-liquid distribution characteristics of the absorber, (2) SO_2 and particulate removal efficiencies as a function of gas velocity and liquid rates, and (3) the effect of spray nozzle type and location and scrubber internals, such as grids and packing, on SO_2 removal.

The gas-liquid distribution study consisted of operating the absorber with (1) air only to determine the gas distribution and (2) air and water to determine the liquid distribution. With the scrubber containing no internals and operating at gas velocities of 12.6 and 19 ft/sec, the airflow was unsymmetrical at the scrubber inlet but became symmetrical at the lower portions of the tower. Five superimposed sections of bar grids (straightening vanes) were installed at the scrubber inlet, which improved the flow profile down the absorber. Waterflow traverses during the air-water tests revealed a poor liquid distribution at all levels. The total flow rate and nozzle pressure had little effect on liquid distribution. The data indicates that a large portion of the waterflow was on the walls at the upper levels but disengaged from the wall as it proceeded down the tower.

Gas velocity profiles were not as good and the liquid distriuution was still poor, after the addition of six inches of Poly Grid packing to each of the five bar grids. However, the packing did tend to cause less flow on the scrubber walls.

Tests were performed using sodium carbonate as the absorbent to determine the maximum efficiency of the scrubber. With no internals but the straightening vanes in the absorber the SO_2 removal ranged from 90 to 97 percent with an average of 94 percent which indicated good mass transfer. The liquid rate had the greatest effect on SO_2 removal; increasing the liquid rate increased SO_2 removal. The spray nozzle type (pressure drop) and its location had a lesser effect; an increase in nozzle pressure drop increased SO_2 removal. A higher SO_2 removal was observed when the total liquid flow was routed through the uppermost nozzles at the scrubber inlet than when the liquid was distributed at various levels down the absorber.

With lime as the absorbent, five 6-inch sections of the Poly Grid packing were added to the bar grids to increase SO_2 removal from the 62 percent obtained with no internals to a prerequisite 85 percent at established conditions (gas velocity of 19 ft/sec, liquid rate of 212 gpm, and liquid to gas ratio of 56). During the test program, the SO_2 removal ranged from 79 to 95 percent and averaged 88 percent. The effect of the liquid rate was more dominant than that of the gas velocity. Increasing the liquid rate was more dominant than that of the gas velocity. Increasing the liquid rate increased SO_2 removal, while decrease in SO_2 removal occurred when the gas velocity was increased. Higher SO_2 removals occurred when the slurry was routed through the top than when it was distributed throughout the tower. The nozzle pressure drop had no significant effect on SO_2 removal. Additional lime tests were made to determine the effect of the number of scrubbing stages (gas-liquid contact time) on SO_2 removal. With the same internals, the scrubbing slurry was successively introduced above each grid down the absorber. Increasing the gas-liquid contact time was found to improve SO_2 removal.

Comparing the liquid distribution data with the SO_2 removal efficiencies for these tests indicates that good SO_2 removals can be achieved with a less than ideal gas-liquid distribution in the absorber. Other scrubber types may also exhibit a similarly poor gas-liquid distribution, but there is little published data for comparison.

When limestone was used as the absorbent, an additional 3 inches of Poly Grid packing (five 9-inch sections) were added to obtain the preferred SO_2 removal efficiency of 85 percent at set conditions. During the test program, the SO_2 removal ranged from 77 to 92 percent with an average of 84 percent. The gas velocity had the greatest effect on SO_2 removal. Increasing the gas velocity from base conditions lowered SO_2 removal efficiencies. An increase in liquid rate increased SO_2 removal. The nozzle pressure drop had no significant effect on SO_2 removal.

The particulate removal averaged 99.4 percent for both the lime and limestone tests.

TVA completed the lime/limestone-scrubbing tests with the 1-MW cocurrent pilot scrubber in mid-1977. These tests successfully demonstrated that the cocurrent scrubber is an effective SO_2 scrubber. The results of these pilot-plant tests were used to guide the design of a 10-MW improved prototype scrubber which was installed in the idle Hydro-Filter scrubber train at the Shawnee facility. Figure 1 is the process flow diagram for the new 10-MW prototype scrubber in the cocurrent mode. The possible advantages of a cocurrent scrubber system over a conventional countercurrent system are:

o The equipment configuration is more compatible with most power plant duct and fan arrangements. The gas would enter the scrubber at a higher elevation and exit near ground level. The entrainment

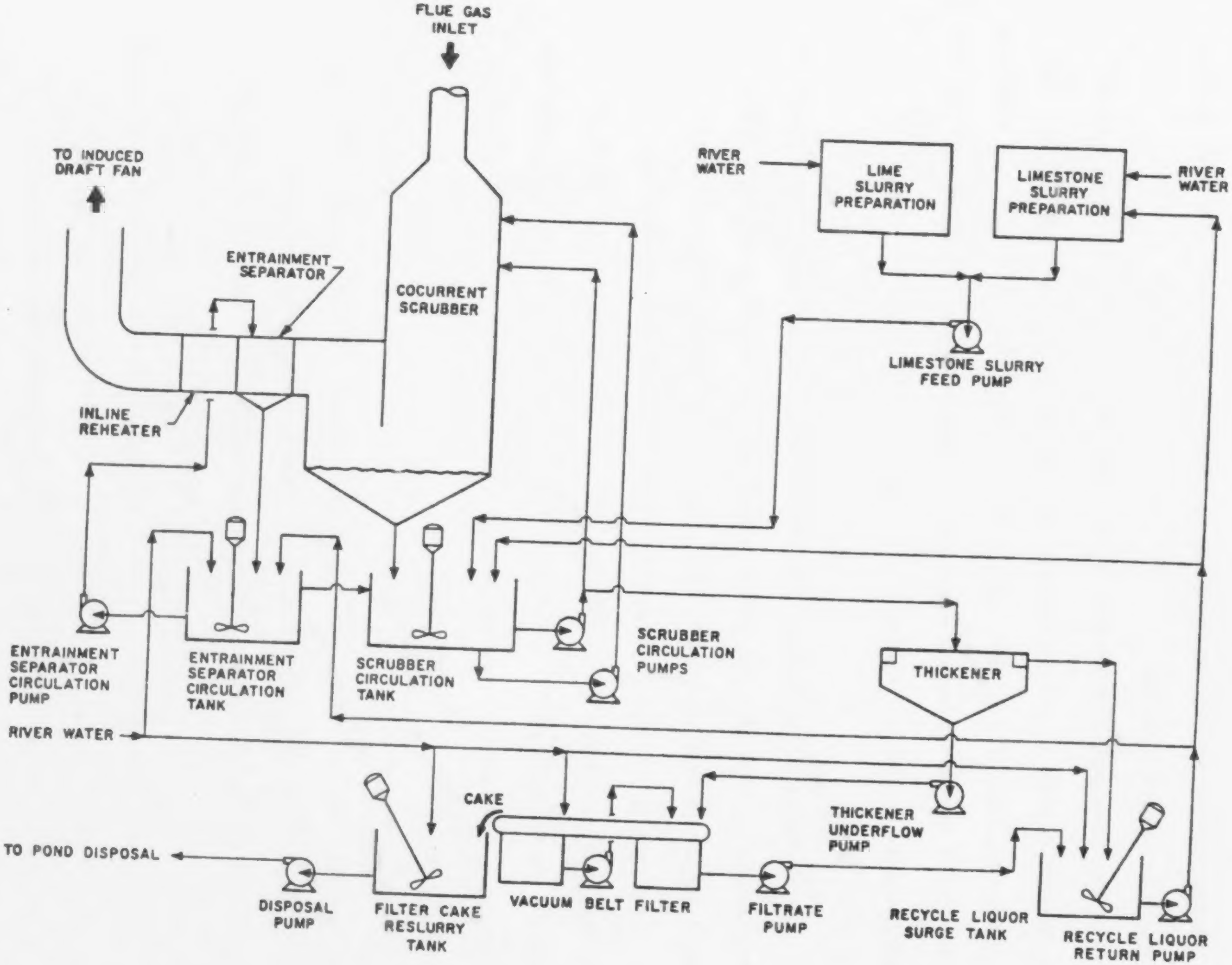

FLUE GAS INLET
TO INDUCED DRAFT FAN
ENTRAINMENT SEPARATOR
COCURRENT SCRUBBER
INLINE REHEATER
RIVER WATER
LIME SLURRY PREPARATION
LIMESTONE SLURRY PREPARATION
RIVER WATER
LIMESTONE SLURRY FEED PUMP
ENTRAINMENT SEPARATOR CIRCULATION PUMP
ENTRAINMENT SEPARATOR CIRCULATION TANK
SCRUBBER CIRCULATION TANK
SCRUBBER CIRCULATION PUMPS
THICKENER
RIVER WATER
CAKE
TO POND DISPOSAL
DISPOSAL PUMP
FILTER CAKE RESLURRY TANK
VACUUM BELT FILTER
FILTRATE PUMP
THICKENER UNDERFLOW PUMP
RECYCLE LIQUOR SURGE TANK
RECYCLE LIQUOR RETURN PUMP

FIGURE I

Process Flow Diagram - Improved Prototype Scrubber (Cocurrent Mode) separator and reheat systems (likely to require the most attention) would be located near ground level. Likewise, the induced-draft (ID) fans would be on the ground and the connecting ductwork to the stack would be shorter and probably less complex.

o The physical arrangement of the proposed cocurrent system causes the gas to change direction in the base of the unit before it enters the entrainment separator. The change in direction, together with the vertical position of the entrainment separator, promotes good liquid separation and drainage. Also, a separate entrainment wash loop can be used if needed.

o Scrubbing liquid would tend to coalesce into larger droplets before it disengaged from the gas stream near the base of the scrubber. This would further facilitate efficient operation of the mist eliminator.

o Flooding of the unit and associated high-pressure loss and excessive entrainment of scrubbing slurry, even if grids are added to improve gas-liquid contact, is less likely. Also, during normal cocurrent operation, the gas-side pressure loss would be lower since some liquid-side energy would be recovered.

o Higher gas velocites (smaller scrubbers) are expected because of the reduced tendency to flood and because more efficient mist elimination is likely. (The Colbert pilot tests were successfully performed at approximately 30 ft/sec superficial gas velocity.) Therefore, smaller or fewer scrubber modules would be required in a full-scale system.

To evaluate further the merits of the cocurrent scrubbing concept and to obtain additional data for scale-up to a full-scale facility, EPRI contracted with TVA to design, procure, erect and test the improved prototype wet-scrubbing system (gas flow equivalent to 10-MW of generating capacity) at the TVA Shawnee Steam Plant, Paducah, Kentucky.

Test Objectives

The objectives of the scrubber test program are as follows:

1. To perform an operating variable study that will identify the optimum operating conditions of the improved prototype scrubber with respect to SO_2 and particulate removal efficiency.
2. To develop the scale-up similarities and differences between the Colbert Test Program (1-MW) and the Shawnee Test Program (10-MW).
3. To determine design parameters for scale-up to a full-scale FGD system.
4. To perform a long-term reliability demonstration of a cocurrent prototype limestone scrubber system at optimum operating conditions.
5. To evaluate the collection efficiency, reliability, and materials of construction of the entrainment separator.
6. To evaluate the performance of an inline-indirect steam reheater with respect to heat transfer characteristics, materials of construction, and reliability.
7. To determine the physical properties of the waste solids which are produced during the reliability demonstration.
8. To evaluate the performance of other mechanical components within the scrubber system, such as pumps, piping materials, valves, and instrumentation.

General Test Program Outline and Schedule

The type of tests and order of testing will be generally as outlined below:

1. Preoperational Testing (Final Equipment Checkout).
2. Sodium carbonate-scrubbing factorial tests to establish the maximum removal efficiency of the scrubber in various operating modes.

3. Limestone-scrubbing operating variable study with a cocurrent scrubber.
    a. Factorial testing.
    b. Short-term reliability tests.
4. Lime-scrubbing operating variable study with a cocurrent scrubber.
    a. Factorial testing.
    b. Short-term reliability tests.
5. Reliability demonstration at optimum operating conditions--cocurrent limestone scrubbing.

A proposed schedule for the test program is shown in Figure 2. The schedule is included to indicate the order of testing and relative time to be assigned to each test block. It will be subject to change as the test program proceeds. The length of individual tests will vary from an 8-hr shift (factorial tests) to a month (reliability demonstration). However, most tests will probably be 1-2 weeks in length to allow the composition of the process liquor and solids to approach equilibrium. TVA technical personnel will determine when the test objective has been achieved and when the tests are to be terminated.

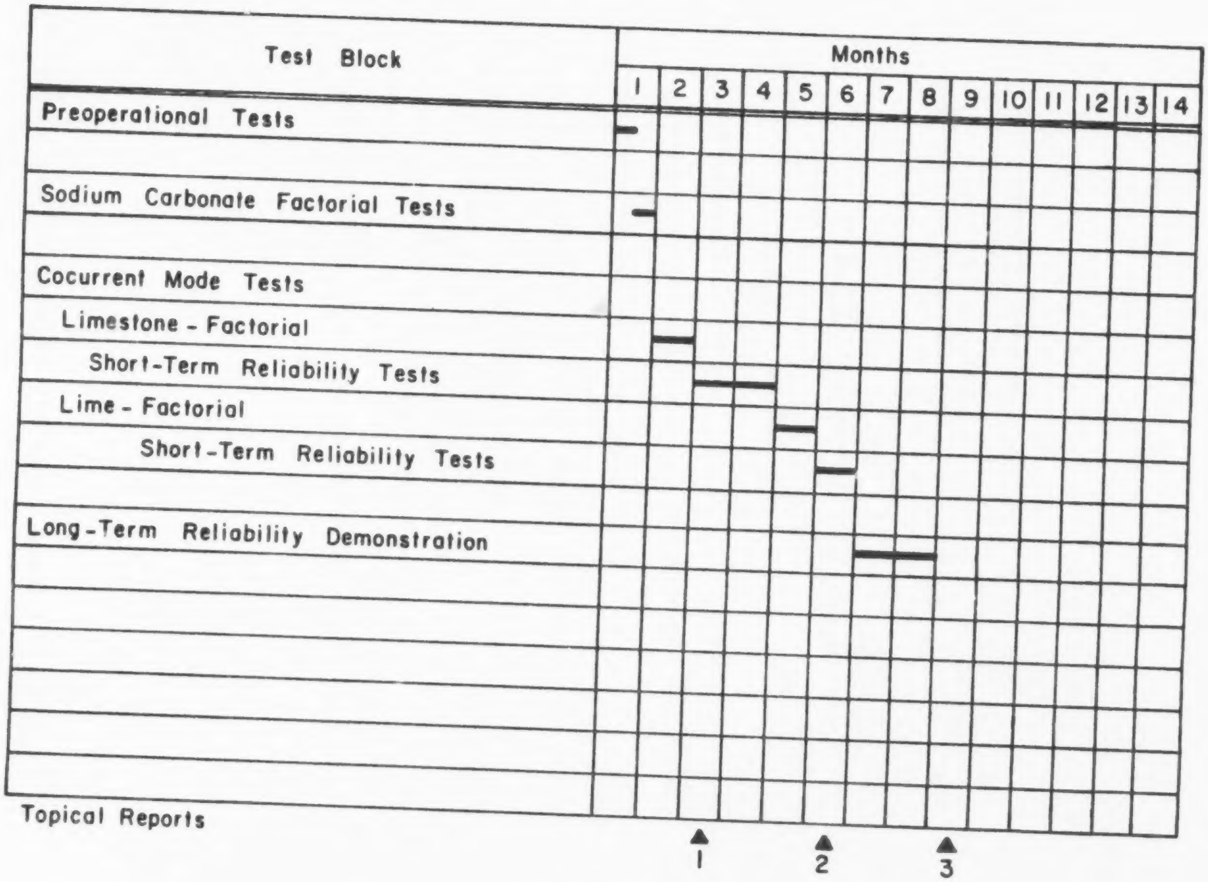
DOWA Process

In this process, SO_2 is absorbed in a clear solution of basic aluminum sulfate. The spent absorbent is oxidized with air, then neutralized with limestone to remove the sulfur in the form of gypsum. The regenerated basic aluminum sulfate solution is recycled to the absorber.

PROCESS DESCRIPTION

The flue gas from the boiler flows directly into the absorber (Figure 3). SO_2 is absorbed by a counterflowing stream of basic aluminum sulfate ($Al_2(SO_4)_3 \cdot Al_2O_3$). The absorption mechanisms are given in

Proposed Test Program Schedule
Improved Prototype Scrubber



1-Report on sodium carbonate tests
2-Report on limestone tests
3-Report on lime tests

Figure 2

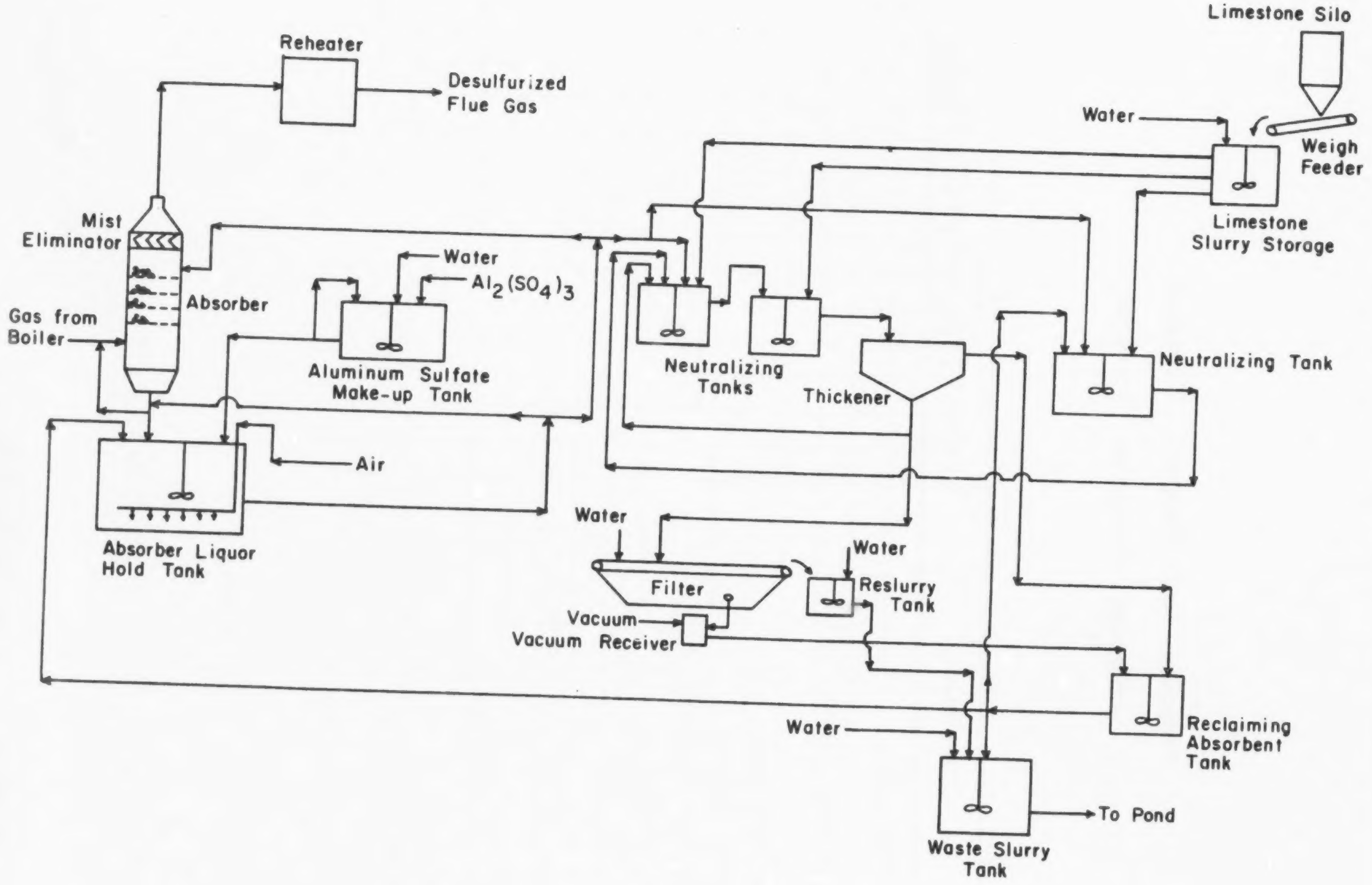

Limestone Silo
Reheater
Desulfurized Flue Gas
Water
Weigh Feeder
Limestone Slurry Storage
Mist Eliminator
Water
$Al_2(SO_4)_3$
Gas from Boiler
Absorber
Aluminum Sulfate Make-up Tank
Neutralizing Tanks
Thickener
Neutralizing Tank
Air
Absorber Liquor Hold Tank
Water
Water
Filter
Reslurry Tank
Vacuum
Vacuum Receiver
Water
Reclaiming Absorbent Tank
Waste Slurry Tank
To Pond

Figure 3. Dowa Typical Process Flow Diagram

Reaction 1 (below). The cleaned gas passes through a mist eliminator before entering the stack.

The spent absorbent is delivered to an oxidizer into which fine air bubbles are injected. The sulfites in the solution are oxidized to sulfates by Reaction 2. The bulk of the resulting solution is recycled to the absorber.

The reminader of the oxidizer effluent is channeled to neutralizing tanks. Limestone ($CaCO_3$) is added to recover the basic aluminum sulfate and precipitate the sulfur in the form of gypsum (Reaction 3). The liquor from the neutralizing tank overflows into a thickener, where the absorbent liquor is separated from the gypsum. Further separation of the gypsum slurry from the thickener takes place in a filter. The resulting gypsum is sent to disposal. The liquor from the filter is mixed with that from the thickener and recycled to the absorber.

REACTIONS

Basic aluminum sulfate is used in the absorber to remove SO_2. The oxidizer is used to convert the resulting sulfites to sulfates. The products are then regenerated in the neutralizer.

Absorber:

$$Al_2(SO_4)_3 \cdot Al_2O_3 + 3SO_2 \longrightarrow Al_2(SO_4)_3 \cdot Al_2(SO_3)_3 \qquad (1)$$

Oxidizer:

$$2Al_2(SO_4)_3\ Al_2(SO_3)_3 + 3O_2 \longrightarrow 4Al_2(SO_4)_3 \qquad (2)$$

Neutralizer:

$$2Al_2(SO_4)_3 + 3CaCO_3 + 6H_2O \longrightarrow \qquad (3)$$

$$Al_2(SO_4)_3 \cdot Al_2O_3 + 3CaSO_4 \cdot 2H_2O + 3CO$$

Advantages

(1) The process is not very complex.

(2) Since limestone is used as a neutralizing agent, operating costs are low.

(3) Due to the low liquid-to-gas ratio, equipment size requirements are low.

(4) Recovered gypsum my be of high quality and, thus, sellable, wherever a demand for it exists. In this project, however, the gypsum will be a disposal byproduct.

DISADVANTAGES/PROBLEMS

(1) Gypsum is not a very desirable commodity for recovery, since other sources can easily supply the existing market.

STATUS

TVA has negotiated with EPA for the use of the TCA for testing of this process.

Initial engineering design and procurement are underway. TVA is negotiating contracts with both UOP and EPRI for the project. An optimistic schedule (Figure 4) calls for modification to the TCA to begin during October with operation of the scrubber to begin in early January.

Test Objectives

The major objectives of this test and demonstration program are as follows:

1. To demonstrate that the DOWA process can effectively treat flue gas from a boiler which is fired with high-sulfur coal and meet current emission standards.

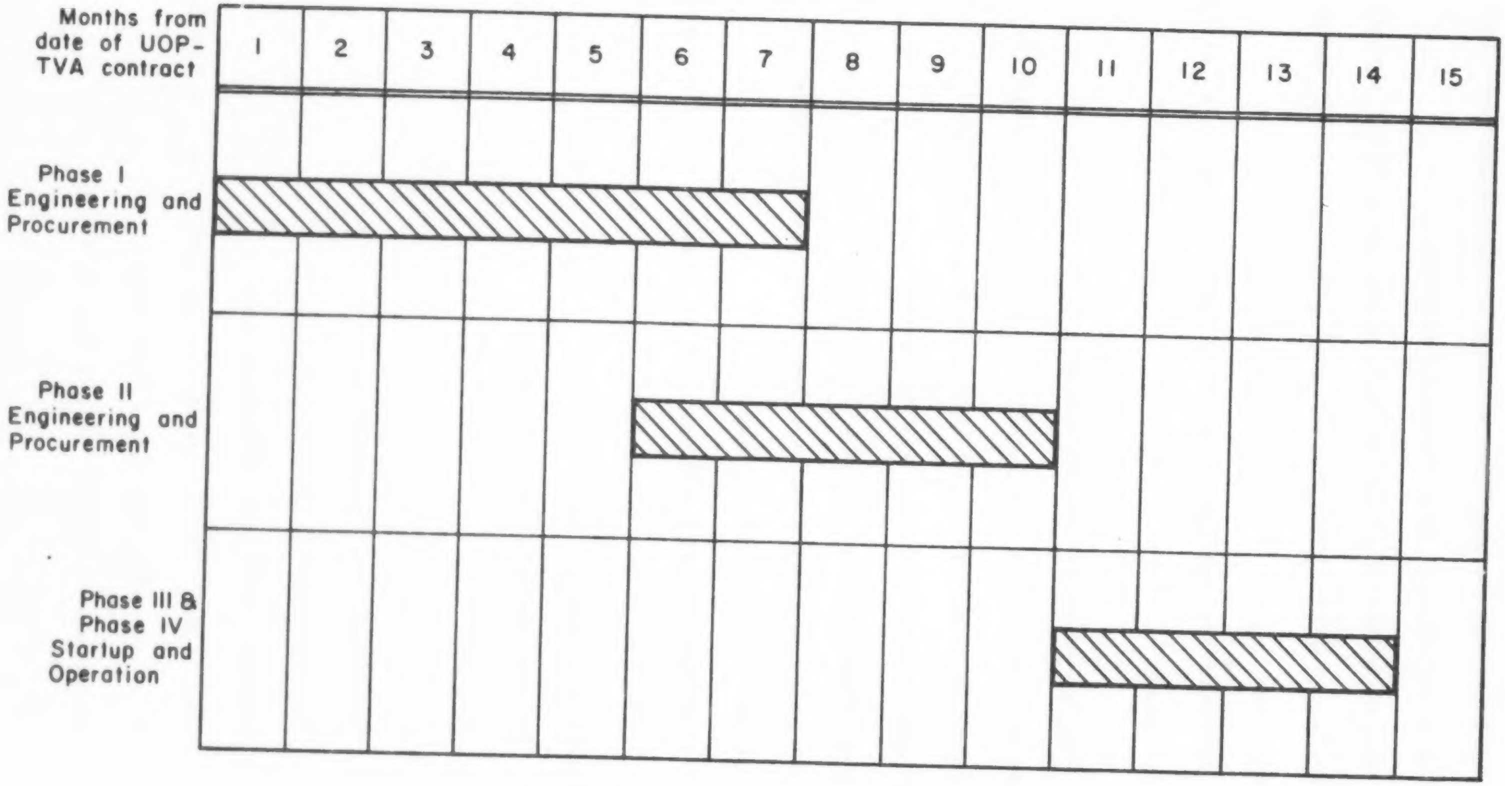


Figure 4. Dowa Project Schedule

2. To evaluate the physical properties of the gypsum byproduct and its suitability as a landfill material.
3. To determine the SO_2 removal efficiency and particulate removal efficiency of the DOWA process over a broad range of operating conditions.
4. To determine or confirm parameters for scale-up to a full-scale FGD system.

General Test Program Description and Schedule

The types of test blocks and operating variables which will be considered for investigation during the test program are outlined below:

1. Equipment shakedown with air and water.
2. System shakedown and process demonstration at operating conditions which are based on previous commercial experience in oil-fired boiler applications.
3. Factorial tests - SO_2 and particulate removal efficiencies; determine the effect of the following variables upon SO_2 and particulate removal efficiencies:
    a. Al concentration
    b. Basicity
    c. TCA pressure drop
    d. TCA liquid recirculation rate
    e. TCA superficial gas velocity
    f. Absorber hold tank retention time
4. Factorial tests - Oxidation; determine the effect of the following variables upon the oxidation efficiency:
    a. Oxygen stoichiometry in the absorber loop

    c. Al concentration
    d. Basicity
    e. Absorber hold tank retention time

5. Factorial tests - Neutralization and byproduct production; determine the effect of the following variables upon neutralization reaction rate, settling rates of precipitates, filterability and final settled bulk density of byproduct gypsum, and aluminum losses.
6. Short-term reliability tests.

Following the system shakedown and initial process demonstration, process variable studies will begin with factorial tests which are designed to screen the effect of each of the above listed variables. If technically feasible and to conserve time, several of the factorial tests may be performed simultaneously. Following the factorial tests, a series of short-term reliability tests will be performed. The selection of operating conditions for these tests will be based upon the results of the factorial tests. The purpose of the short-term reliability tests is to obtain more definitive information on the effect of operating conditions upon the reliability of system components and to obtain operating data which will be the basis for selection of operating conditions for a long-term reliability demonstration. The conditions of the scrubber equipment will be evaluated at the end of each test. Although the operating conditions for all tests will be specified before each test block begins, the test conditions will be subject to change as the results of the initial tests are evaluated.

If results from this test program are favorable, additional funding will be needed for more extensive variable studies and a long-term reliability demonstration of the DOWA FGD system. As indicated above, the selection of the operating conditions for the long-term reliability test should be based on the short-term reliability test results. System reliability and economy of operation will be the major criteria which should be used in the selection of operating conditions. Also, only operating conditions which have met the current emission standards should be selected. During the reliability test, all the scrubber operating conditions should be held constant except for the flue gas rate which will be varied in proportion to the boiler load. During this test the physical and chemical properties of the gypsum byproduct should be routinely determined, including settling rate, final settled bulk density, compressive strength, filterability, particle size distribution, chemical analysis, and general crystal form. To the extent possible, pertinent physical properties should be correlated with the operating conditions of the process unit.

The preoperational tests (equipment shakedown) and the test program will be conducted during a 4-month period. A test program schedule which indicates the order of tests and the relative amount of time assigned to each test will be proposed to EPRI and UOP one month prior to startup. The test program schedule will be subject to revision as the test results are evaluated. All schedule changes will be approved by EPRI and UOP.

DEMONSTRATION OF FORCED OXIDATION AT WIDOWS CREEK UNIT 8

The purpose of oxidizing lime/limestone FGD product sludges is to convert calcium sulfite ($CaSO_3$), the normal product of these scrubber processes to calcium sulfate ($CaSO_4$). Calcium sulfate, which is commonly known as gypsum, is a more desirable waste product because it improves the settling and dewatering properties of the sludge. This in turn reduces the volume of material for disposal and makes a material which may be suitable for landfill without need of additives or the use of mixing or blending equipment.

This project is designed to develop, demonstrate, and evaluate, at the full-scale level, the technical feasibility and environmental acceptability of utilizing oxidation as a method of sludge disposal. This will be accomplished by testing a forced oxidation system on the "D" train (140-MW equivalent) of the Widows Creek unit 8 wet limestone scrubber. A slipstream of 25 percent of the full load flow of the effluent slurry will be treated in a thickener and vacuum filter.

Tests will include two-stage forced oxidation, single-stage forced oxidation, and oxidation in both stages (venturi and absorber). The effect of such variables as air stoichiometry, pH, and limestone stoichiometry will also be evaluated.

Combustion Engineering, Inc., under a contract with the Tennessee Valley Authority will be responsible for the design, procurement, erection, testing, and reporting of this oxidation demonstration program.

A. Widows Creek Simulation at Shawnee

Essentially complete oxidation of calcium sulfites from the scrubbers operated at the Shawnee Test Facility has been routinely achieved during forced oxidation testing on both the TCA and the venturi/spray tower systems. However, in pilot-plant scale tests conducted at the TVA Colbert plant, the sludge could not be oxidized in any practical manner. These opposed results indicate some undetermined site-specific conditions that either promotes oxidation at Shawnee or conversely deters oxidation at Colbert. Since the natural oxidation at Widows Creek and Colbert is essentially the same and is significantly lower than that obtained at Shawnee, considerable concern has arisen over the Colbert-Shawnee correlation since extrapolation may be possible to the forced oxidation mode. Should this be the case, little to no oxidation could be achieved on the Widows Creek system if a forced oxidation should be desired for use as a method of treating the sludge. In order to predict achievable oxidation at Widows Creek, a special coal burn at Shawnee using Widows Creek coal, scrubbing with Widows Creek limestone, and using the venturi-spray tower oxidation mode as operated in the EPA test program is planned. The main problem would be the difference in boilers for the two systems and the possible effect this could have on oxidation; Shawnee--B&W, Widows Creek unit 8--CE. These tests would simulate the Widows Creek scrubber design as much as possible so that problem areas could be identified.

B. Widows Creek Forced Oxidation

TVA will also evaluate oxidation on one of the scrubber trains at the Widows Creek unit 8 facility such that data can be collected on the feasability of oxidizing the sludge under Widows Creek operating condition. Figure 5 is a flow diagram of the Widows Creek Forced Oxidation System. Should oxidation prove to be technically feasible an economic evaluation will be made for comparison to other methods of disposal, i.e., sludge-fly ash blending. This project is designed to develop,

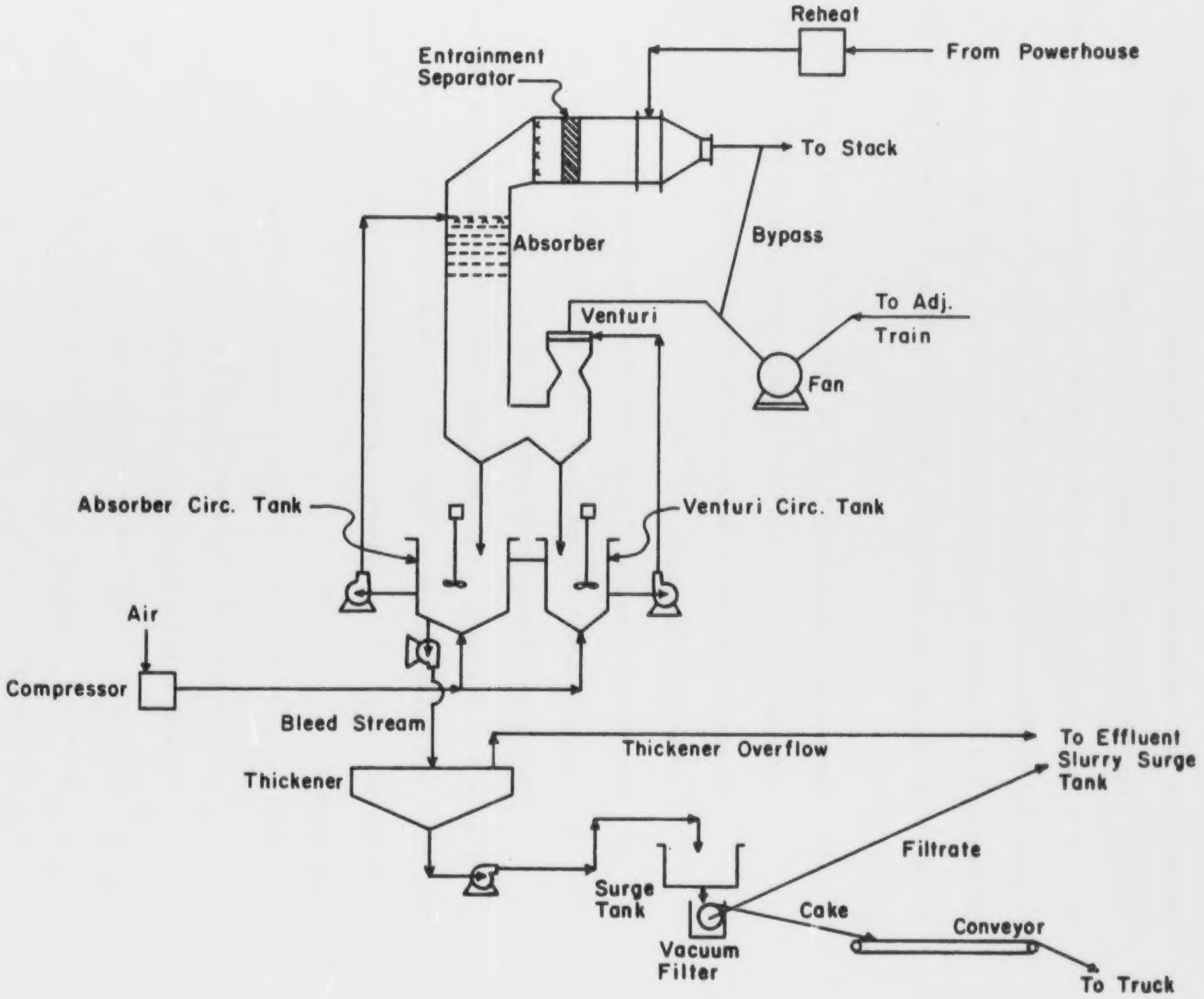

Reheat
From Powerhouse
Entrainment Separator
To Stack
Absorber
Bypass
Venturi
To Adj. Train
Fan
Absorber Circ. Tank
Venturi Circ. Tank
Air
Compressor
Bleed Stream
Thickener Overflow
To Effluent Slurry Surge Tank
Thickener
Filtrate
Surge Tank
Cake
Conveyor
Vacuum Filter
To Truck

Figure 5. Flow Diagram - Widows Creek Forced Oxidation

demonstrate, and evaluate the technical feasibility and environmental acceptability of oxidation as a method for disposal of sludge from the No. 7 and No. 8 wet/limestone scrubber units at Widows Creek. The primary purpose for oxidizing the scrubber solids is to improve waste solids dewatering and landfill disposal characteristics. This objective will be accomplished through specific tests which will be made to meet the following criteria.

1. The project should result in an evaluation that will provide a reference point for making confident decisions on the feasibility of converting the scrubbers at the Widows Creek station to the forced oxidation mode.
2. Provide the results of this evaluation by March 1979 so that a sludge disposal system can be installed and be ready for use before the Widows Creek pond is filled.

Figure 6 is a schedule for the Widows Creek Forced Oxidation System. This study will also include a comparative economic study of forced oxidation vs. other sludge disposal processes as they relate to the Widows Creek system.

Other major factors that will be included for evaluation are:

1. Transportability of the oxidized solids after dewatering.
2. Primary application for the dewatered solids.
3. Material handling equipment required.
4. Methods of disposal or reuse of waste water resulting from dewatering.
5. Effect of dewatering for landfill treatment.

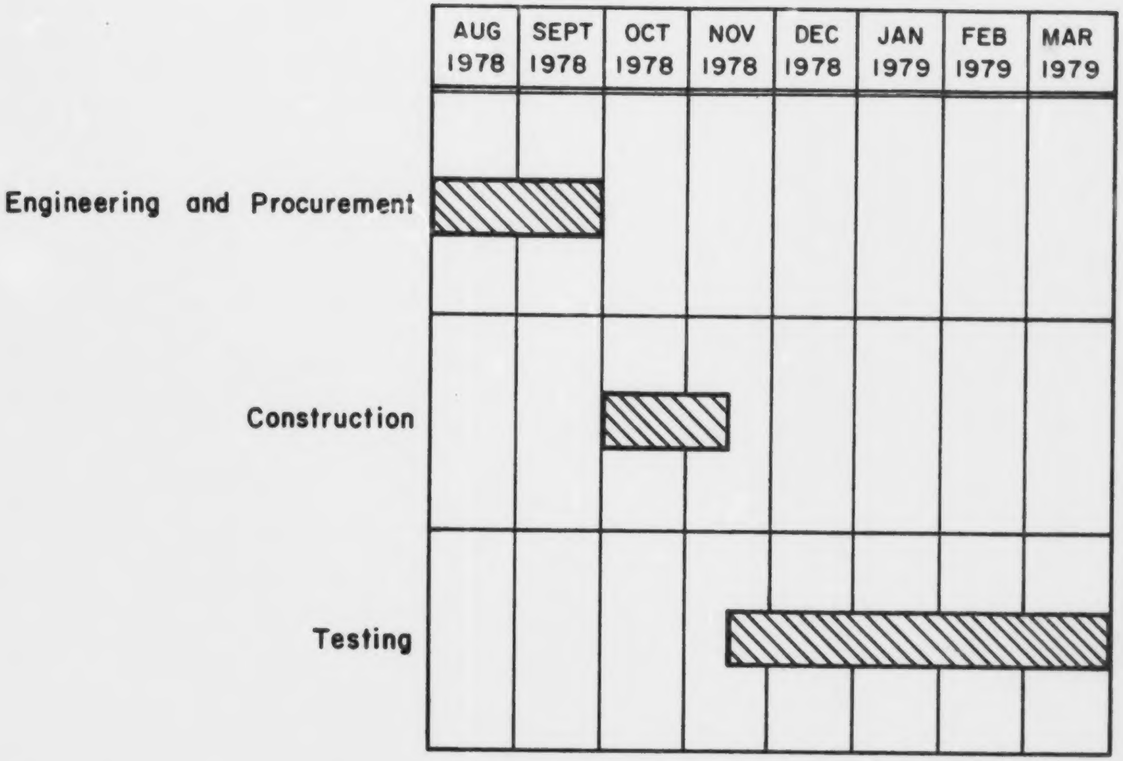


Figure 6. Widows Creek Forced Oxidation Schedule

CURRENT STATUS OF DEVELOPMENT OF THE SHAWNEE LIME-LIMESTONE COMPUTER PROGRAM

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Sponsored by the U.S. Environmental Protection Agency
Royal Villa Motel in Raleigh, North Carolina
August 29, 1978

CURRENT STATUS OF DEVELOPMENT OF THE SHAWNEE LIME-LIMESTONE COMPUTER PROGRAM

GENERAL SCOPE AND PURPOSE

In conjunction with the U.S. Environmental Protection Agency (EPA)-sponsored Shawnee test program, Bechtel National, Inc., and the Tennessee Valley Authority (TVA) have jointly developed a computer program capable of projecting comparative investment and revenue requirements for lime and limestone scrubbing systems. The computer program has been developed to permit the estimation of relative economics of these systems for variations in process design alternatives (i.e., limestone vs lime scrubbing, alternative scrubber types, or alternative sludge disposal methods) or variations in the values of independent design parameters (i.e., scrubber gas velocity and L/G ratio, alkali stoichiometry, slurry residence time, reheat temperature, and specific sludge disposal design). Although the program is not intended to compute the economics of an individual system to a high degree of accuracy, it is based on sufficient detail to allow the quick projection of preliminary conceptual design and costs for various lime-limestone case variations on a common design and cost basis.

PROGRAM DEVELOPMENT

The responsibility in the development of the computer program was shared by Bechtel and TVA. Bechtel's major responsibility was to analyze the results of the Shawnee scrubbing tests and develop models for calculating the overall material balance flow rates and stream compositions. Bechtel provided TVA with a complete computer program for specifying this information. TVA was responsible for determining the size limitations of the required equipment for establishing the minimum number of parallel equipment trains, accumulating cost data for the major equipment items, and developing models for projecting equipment and field material costs as a function of equipment capacity. Utilizing these relationships TVA developed models to project the overall investment cost breakdown and a procedure for using the output of the material balance and investment models as input to a previously developed TVA program for projecting annual and lifetime revenue requirements.

PROGRAM DESCRIPTION AND DOCUMENTATION

TVA has presented two papers which reported the status of the program development and displayed its capabilities. The first, titled "Shawnee Limestone-Lime Scrubbing Process Computerized Design-Cost Estimate Program: Summary Description Report," was given at the Industry Briefing Conference sponsored by EPA at Raleigh, North Carolina, October 14, 1976. The second, titled "Economic Evaluation Techniques, Results, and Computer Modeling for Flue Gas Desulfurization," was presented at the EPA Flue Gas Desulfurization Symposium in Hollywood, Florida, November 8-11, 1977.

A significant number of revisions have been incorporated into the program since these earlier reports. The present paper describes the current capabilities of the computer program. Since the design basis for the lime and limestone scrubbing systems has not changed appreciably from the earlier publications, it will not be included in this paper. TVA is currently in the process of preparing a users manual for the overall program which will include all the information required for running the Turbulent Contact Absorber (TCA) program option. It is anticipated that the users manual will be available within the next few months.

CURRENT PROGRAM SCOPE

Uses and Limitations

The present computer program has the capability of projecting a complete conceptual design package for lime or limestone scrubbing utilizing a TCA with any one of four sludge disposal options (discussed later). The program is designed to consider new coal-fired power units ranging in size from 100-1300 MW. Equipment size and layout configurations are modeled based on coals ranging in sulfur contents from 2-5%. To limit the extremely wide variations in equipment sizes and layout configurations which can result with changes in other key independent variables, the following range of values for these variables was established.

Scrubber gas velocity	8-12.5 ft/sec
Liquor recirculation rate	25-75 gal/kft^3
Slurry residence time in hold tank	2-25 min

However, operating parameters and plant sizes outside of these ranges will not necessarily be invalid.

It is expected that the results may within limits also be valid for extrapolation of sulfur content of coal beyond the range actually tested at Shawnee. The Shawnee models are based on scrubbing results over an SO_2 concentration range of approximately 1500 to 4000 ppm.

The effect of variations in any of the inputs, such as scrubber gas velocity, degree of SO_2 removal, reheat temperature, alkali stoichiometry, L/G ratio, etc., on process design and economics may be determined.

For both lime and limestone scrubbing, SO_2 removal, stoichiometry (pH for the lime option), and L/G ratio may all be specified and results projected. Alternatively, SO_2 removal and stoichiometry (or pH) may be specified and L/G calculated, or SO_2 removal and L/G ratio may be specified and stoichiometry (or pH) calculated. An additional option is being incorporated into the program allowing for the calculation of SO_2 removal based on input values of L/G and stoichiometry (or pH).

The outputs of the overall computer program include (1) a detailed material balance including properties of the major streams, (2) a detailed water balance itemizing water availability and water required, (3) specifications of the scrubber system design, (4) a revised method for displaying overall pond design and costs, (5) specifications and costs of the process equipment by major processing area, (6) a detailed breakdown of the projected capital investment requirements, (7) an itemized breakdown of the projected revenue requirements by component for the first year of operation of the system, and (8) a lifetime revenue requirement analyses showing projected costs for each year of operation of the plant, as well as lifetime cumulative and discounted costs and equivalent unit revenue requirements.

New Program Options and Modifications Since the November 1977 FGD Symposium

To illustrate the current program inputs and outputs, an example run of the updated computer program is shown in the appendix for limestone scrubbing with an onsite pond disposal option. Discussions of the modifications incorporated into the program since the November 1977 FGD symposium are given below.

Particulate Removal--

The quantity of ash in the coal which is emitted overhead as fly ash is an input. Additional inputs are required to specify ash removal upstream of the scrubber and within the scrubber. These are input either as a removal efficiency in percent or as an equivalent outlet emission in pounds per million Btu. If removal is not input, a 33% efficient mechanical collector is provided for protecting the fan from abrasion by large fly ash particles. A cost model is available for optionally including the costs for the mechanical collector. As discussed later, cost models for a high-efficiency electrostatic precipitator (ESP) or baghouse are presently being incorporated into the program, but are not yet available for the current version. An example output of the fly ash removal option is shown on pages 10 and 13 of the appendix under the heading "Fly Ash Removal."

SO_2 Removal--

The percentage of the sulfur in coal which is emitted overhead as SO_2 is an input to the program. The degree of SO_2 removal may be input by specifying either

1. % SO_2 removed,
2. lbs SO_2 emitted per million Btu of heat input to the boiler, or
3. ppm SO_2 in outlet flue gas.

For each computer run, equivalent SO_2 emissions are displayed on all three basis, regardless of the method for inputting the degree of SO_2 removal. The alternative methods for specifying SO_2 removal are illustrated on page 14 under the heading "Flue Gas to Stack."

Redundancy--

In addition to designing the flue gas desulfurization (FGD) system with spare operating pumps and optionally with spare scrubbing trains as described in the earlier publications, spare feed preparation units may now be specified. This is applicable for both limestone scrubbing in which spare ball mill trains can be provided, and lime scrubbing in which spare slakers can be specified. The number of redundant alkali preparation units and redundant scrubber trains are specified under "Raw Material Handling Area" (pp. 11 and 20) and "Scrubber System Variables" (pp. 11, 16, and 22) respectively.

Water Balance--

The program has previously assumed no net accumulation or loss of water due to rainfall, evaporation, and seepage. The current version of the program allows for specific rainfall, evaporation, and seepage rates to be input for accurately projecting makeup water requirements. The water balance inputs and outputs for an example run are shown on page 15.

Waste Disposal Options--

The program allows for four alternate waste disposal alternatives to be assessed including:

1. Onsite ponding
    a. Unlined pond
    b. Clay-lined pond (cost of clay and depth of lining input)
    c. Synthetic-lined pond (cost of liner input)
2. Thickener - ponding
3. Thickener - fixation fee
4. Thickener - filter - fixation fee

The onsite ponding options may also be run with fixation fees applied to them. For alternatives 3 and 4, the fixation fee must include costs for transportation and disposal of the fixed sludge offsite. For alternatives 1 and 2, however, only the costs for fixation need be provided since the fixed sludge can be disposed of at the existing pond site.

For the ponding alternatives the program allows for the onsite pond to be sized larger or smaller than the normal projected lifetime capacity.

This option has been incorporated (1) to account for variations in the sulfur content of fuel, (2) to evaluate design philosophy in construction ponds for less than the total amount of sludge to be disposed of (requires assessment of additional cost for expanding pond later), or (3) to allow the feed preparation and scrubbing areas to be sized based on maximum sulfur contents expected, while sizing the pond based on average sulfur contents. An example output of the revised onsite ponding model is shown on page 19.

Equipment Cost Breakdown--

The program has been modified to provide a breakdown showing projected equipment specifications and cost for each equipment item. Both material and labor components of equipment costs are displayed for each of the three major areas. The equipment list and costs for limestone scrubbing and onsite ponding are illustrated on pages 20-23. Although a complete printout for the lime and alternate sludge disposal options is not included in this paper, an equipment list for those options is illustrated on pages 27-30.

Operating Profile--

The current version of the program allows for the specification of three alternative operating profiles as indicated below for projecting lifetime revenue requirements:

1. Profile similar to that utilized in the report Detailed Cost Estimates for Advanced Effluent Desulfurization Processes by G. G. McGlamery, et al. (EPA-600/2-75-006, January 1975)
2. Historical power plant operating profile based on FPC Form 67 data
3. Variable profile with annual load factors as input

See pages 31-32 for illustration.

FUTURE PROGRAM DEVELOPMENT

Further additions to the program are expected to be made as additional test data from Shawnee become available. Bechtel and TVA are currently incorporating the results of the venturi-spray tower tests at Shawnee into a design and cost model for that option. In addition, cost models for upstream fly ash removal by hot or cold side ESP, baghouse collectors and venturi scrubbers are being incorporated into the program. Other options which are being considered for incorporation as sufficient data become available include (1) series scrubbers/high alkali utilization systems and (2) forced oxidation systems.

PROJECTED PROGRAM USE

Upon completion of the overall effort, the program will be useful for projecting a complete conceptual design package for lime-limestone scrubbing including material balance, capital investment estimate, and projected revenue requirements. It is expected that the program will be used by utility companies and architectural and engineering contractors involved in the selection and design of SO_2 removal facilities for specific applications. It is not intended to be used for projecting a final design of a given system, but to assist in the evaluation of system alternatives prior to development of a detailed design. Also, the program will be useful for evaluating the potential impact of various process variables on economics as a guide for planning research and development activities.

Although the program was not meant to be used for comparing projected lime-limestone economics with economics for alternate processes, these comparisons should be valid as long as the basis for the alternate process economics are comparable to those included in the computer program for lime-limestone systems.

Method for Attaining Results

TVA is in the process of loading the current version of the program on a Control Data Corporation (CDC) timesharing computer system and publishing a users manual for utilizing the program. After this effort is complete, outside users will be allowed to access the computer program for making computer runs. Until these activities are completed, TVA, under a Technology Transfer Contract with EPA, can upon request make computer runs for interested users, or can release copies of the program to interested users along with available documentation for running it.

Current Users and Program Applications

A significant number of responses have already been handled under the above arrangement.

Given below is a list of people who have been provided tape copies of the Shawnee lime-limestone computer program.

Robert H. Boeckmann, Gibbs and Hill, Inc.
Paul S. Farber, Argonne National Laboratory
D. J. Hagerty, University of Louisville
J. Scott Hartman, PEDCo
R. G. Knight, Michael Baker, Jr., Inc.
M. Lieberstein, The City of New York,
 Environmental Protection Administration
F. Y. Murad, Combustion Equipment Associates, Inc.
Edward S. Rubin, Carnegie-Mellon University
J. G. Stevens, Exxon
John Valente, Air Correction Division, UOP
John Wysocki, Burns and Roe, Inc.

A list of people who have been provided results of specific computer runs is shown below:

Randy M. Cole, Tennessee Valley Authority
Wilson Cramer, U.S. Steel Corporation
Richard Furman, Florida Power and Light
Robert Gosik, Environmental Protection Agency (Denver)
Robert Lane, Illinois Commerce Commission
S. J. Lutz, TRW, Inc.
M. F. Patterson, Linde Division, Union Carbide Corporation
A. V. Slack, SAS Corporation
John Wile, National Economic Research Associates, Inc.

To date several uses of the program other than those for which it was intended have been tried. The program has been run simulating both industrial and utility boilers, smelter situations, partial scrubbing situations, plant optimization studies, and for comparisons of front end coal cleaning economics with total scrubbing. Probably the most important use to date has been support work for the National Economic Research Association's assessment of the impact of possible NSPS revisions on the electric utilities industry on a national scale. [EPA issued Federal Standards of Performance for New Stationary Sources (often called "new source performance standards" or NSPS).] Examples of some of the sensitivities which might be assessed are shown on pages 33-46.

APPENDIX

INPUT DATA FOR EXAMPLE

PROGRAM RUN

```
1 1 1 1 1
1 1 1 1 1 1 1 1 1 0 1 1 0 0 0
1 1 1
INDUSTRY BRIEFING-500 MW
0 500 9000 10500 33 300 2 175 470 751
57.56 4.14 7.00 1.29 3.12 0.15 16.00 10.74 95 80 1 97.5 80
50 20 12.5 25 2 1.2 12 1 0.0 1 0 0 2.85 5 0 0
15 40 .2 40 30 60 1.2 7.0 1 100 0
2 3 4 5 51 .000001 42 10 1.35 1 4 2 .1
1 12 9999 3500 25 25 5280 1 10 2.5
9 16 5 10 8 12 11.6 8 3 50 10 17.2 1.17
8 50 12.5 2.0 0.12 0.029 17 220.9 178.2 1977 1978
1 1
END
```

OUTPUT OF EXAMPLE RUN

```
                    *** INPUTS ***


BOILER CHARACTERISTICS
------ ---------------

MEGAWATTS =   500.

BOILER HEAT RATE =  9000. BTU/KWH

EXCESS AIR =  33. PERCENT, INCLUDING LEAKAGE

HOT GAS TEMPERATURE = 300. DEG F

COAL ANALYSIS, WT % AS FIRED :

 C       H       O       N       S       CL     ASH     H2O
57.56   4.14    7.00    1.29    3.12    0.15   16.00   10.74

SULFUR OVERHEAD =  95.0 PERCENT

ASH OVERHEAD =  80.0 PERCENT

HEATING VALUE OF COAL = 10500. BTU/LB

                              EFFICIENCY,     EMISSION,
FLYASH REMOVAL                    %           LBS/M BTU
------ -------               -----------     ---------

UPSTREAM OF SCRUBBER             97.5            0.30

WITHIN SCRUBBER                  80.0            0.06

COST OF UPSTREAM FLYASH REMOVAL EXCLUDED

ALKALI
------

LIMESTONE :

          CACO3          =  97.15 WT % DRY BASIS
          SOLUBLE MGO =   0.0
          INERTS         =   2.85
          MOISTURE CONTENT =    5.00 LB H2O/100 LBS DRY LIMESTONE
          LIMESTONE HARDNESS WORK INDEX FACTOR = 10.00
          LIMESTONE DEGREE OF GRIND FACTOR =  1.35

FLY ASH :

          SOLUBLE CAO =    0.0   WT %
          SOLUBLE MGO =    0.0
          INERTS       = 100.00
```

```
RAW MATERIAL HANDLING AREA
--- -------- -------- ----

NUMBER OF REDUNDANT ALKALI PREPARATION UNITS =     1

SCRUBBER SYSTEM VARIABLES
-------- ------ ---------
NUMBER OF OPERATING SCRUBBING TRAINS =   4

NUMBER OF REDUNDANT SCRUBBING TRAINS =   2

NUMBER OF BEDS =   3

NUMBER OF GRIDS =   4

HEIGHT OF SPHERES PER BED =  5.0 INCHES

LIQUID-TO-GAS RATIO =  55. GAL/1000 ACF

SCRUBBER GAS VELOCITY = 12.5 FT/SEC

SO2 EMISSION LIMIT = 1.20 LB SO2/M BTU

STOICHIOMETRY RATIO      TO BE CALCULATED

ENTRAINMENT LEVEL = 0.10 WT %

EHT RESIDENCE TIME =  12.0 MIN

SO2 OXIDIZED IN SYSTEM =  30.0 PERCENT

SOLIDS IN RECIRCULATED SLURRY =  15.0 WT %

SOLIDS DISPOSAL SYSTEM
------ -------- ------

COST OF LAND =  3500.00 DOLLARS/ACRE

SOLIDS IN SYSTEM SLUDGE DISCHARGE =  40.0 WT %

MAXIMUM POND AREA =  9999. ACRES

MAXIMUM EXCAVATION =  25.00 FT

DISTANCE TO POND =  5280. FT

POND LINED WITH 10.0 INCHES CLAY
```

```
STEAM REHEATER (IN-LINE)
----- -------- ---------

SATURATED STEAM TEMPERATURE =  470. DEG F

HEAT OF VAPORIZATION OF STEAM =  751. BTU/LB

OUTLET FLUE GAS TEMPERATURE = 175. DEG F

SUPERFICIAL GAS VELOCITY (FACE VELOCITY) =  25.0 FT/SEC


  IT      SR        SROLD
   1      1.24       1.50
   2      1.24       1.24
   3      1.24       1.24
PAPER PRINT 01,  POND  FEE
```

```
                    *** OUTPUTS ***

HOT GAS TO SCRUBBER
--- --- -- --------

          MOLE PERCENT      LB-MOLE/HR      LB/HR

CO2          12.315         0.2054E+05     0.9040E+06
HCL           0.011         0.1813E+02     0.6612E+03
SO2           0.238         0.3962E+03     0.2538E+05
O2            4.827         0.8050E+04     0.2576E+06
N2           73.867         0.1232E+06     0.3452E+07
H2O           8.743         0.1458E+05     0.2627E+06

SO2 CONCENTRATION IN SCRUBBER INLET GAS = 2376. PPM

FLYASH EMISSION =  0.30 LBS/MILLION BTU
                =  0.152 GRAINS/SCF (WET)   OR   1371. LB/HR

          SOLUBLE CAO IN FLY ASH =      0. LB/HR
          SOLUBLE MGO IN FLY ASH =      0.

HOT GAS FLOW RATE =  .1054E+07 SCFM (60 DEG F, 1 ATM)
                  =  .1540E+07 ACFM (300. DEG F, 1 ATM)

CORRESPONDING COAL FIRING RATE =  .4286E+06 LB/HR

HOT GAS HUMIDITY =  0.057 LB H2O/LB DRY GAS

WET BULB TEMPERATURE = 127. DEG F


WET GAS FROM SCRUBBER
--- --- ---- --------

          MOLE PERCENT      LB-MOLE/HR      LB/HR

CO2          11.683         0.2087E+05     0.9185E+06
SO2           0.047         0.8430E+02     0.5400E+04
O2            4.480         0.8003E+04     0.2561E+06
N2           68.969         0.1232E+06     0.3452E+07
H2O          14.820         0.2647E+05     0.4769E+06

SO2 CONCENTRATION IN SCRUBBER OUTLET GAS =  472. PPM

FLYASH EMISSION =  0.06 LBS/MILLION BTU
                =  0.028 GRAINS/SCF (WET)   OR    274. LB/HR

TOTAL WATER PICKUP =  439.   GPM
           INCLUDING   10.2 GPM ENTRAINMENT

WET GAS FLOW RATE =  .1128E+07 SCFM (60 DEG F, 1 ATM)
                  =  .1274E+07 ACFM (127. DEG F, 1 ATM)

WET GAS SATURATION HUMIDITY =  0.103 LB H2O/LB DRY GAS
```

```
FLUE GAS TO STACK
---- --- -- -----

              MOLE PERCENT      LB-MOLE/HR      LB/HR

CO2              11.665         0.2087E+05      0.9185E+06
SO2               0.047         0.8430E+02      0.5400E+04
O2                4.473         0.8003E+04      0.2561E+06
N2               68.860         0.1232E+06      0.3452E+07
H2O              14.955         0.2676E+05      0.4820E+06

CALCULATED SO2 REMOVAL EFFICIENCY =  78.8 %

SPECIFIED  SO2 EMISSION =    1.20 POUNDS PER MILLION BTU

CALCULATED SO2 CONCENTRATION IN STACK GAS =     471. PPM

FLYASH EMISSION =  0.06 LBS/MILLION BTU
                =  0.028 GRAINS/SCF (WET)  OR    274. LB/HR

STACK GAS FLOW RATE =  .1130E+07 SCFM (60 DEG F, 1 ATM)
                    =  .1380E+07 ACFM (175. DEG F, 1 ATM)


STEAM REHEATER (IN-LINE)
----- -------- ---------

SUPERFICIAL GAS VELOCITY (FACE VELOCITY) =  25.0 FT/SEC

SQUARE PIPE PITCH = 2 TIMES ACTUAL PIPE O.D.

SATURATED STEAM TEMPERATURE =  470. DEG F

OUTLET FLUE GAS TEMPERATURE = 175. DEG F

REQUIRED HEAT INPUT TO REHEATER = 0.6883E+08 BTU/HR

STEAM CONSUMPTION = 0.9165E+05 LBS/HR


                                      HEAT TRANSFER
OUTSIDE PIPE    PRESSURE DROP,        COEFFICIENT,
DIAMETER, IN.   IN. H2O               BTU/HR FT2 DEG F
   1.00              0.76                0.2082E+02

                REHEATER            NUMBER OF
                OUTSIDE PIPE        PIPES PER        NUMBER OF
                AREA, SQ FT         BANK PER         BANKS (ROWS)
                PER TRAIN           TRAIN            PER TRAIN

INCONEL         0.1284E+04             87                3
CORTEN          0.1313E+04             87                4
TOTAL           0.2597E+04             87                7
```

```
ATER BALANCE INPUTS
---- ------- ------


  RAINFALL(IN/YEAR)                        51.
  POND SEEPAGE(CM/SEC)*10**8              100.
  POND EVAPORATION(IN/YEAR)                42.


ATER BALANCE OUTPUTS
---- ------- -------


ATER AVAILABLE

  RAINFALL                                658. GPM                   328778. LB/HR
  ALKALI                                    4. GPM                     1995. LB/HR
       TOTAL                              662. GPM                   330773. LB/HR


ATER REQUIRED

  HUMIDIFICATION                          429. GPM                   214217. LB/HR
  ENTRAINMENT                              10. GPM                     5103. LB/HR
  DISPOSAL WATER                          157. GPM                    78373. LB/HR
  HYDRATION WATER                          11. GPM                     5307. LB/HR
  CLARIFIER EVAPORATION                     0. GPM                        0. LB/HR
  POND EVAPORATION                        575. GPM                   287557. LB/HR
  SEEPAGE                                 160. GPM                    80040. LB/HR

  TOTAL WATER REQUIRED                   1342. GPM                   670596. LB/HR

ET WATER REQUIRED                         680. GPM                   339823. LB/HR
```

```
SCRUBBER SYSTEM
-------- ------

TOTAL NUMBER OF SCRUBBING TRAINS (OPERATING+REDUNDANT) =  6

SO2 REMOVAL =  78.7 PERCENT

PARTICULATE REMOVAL IN SCRUBBER SYSTEM =  80.0 PERCENT

TCA PRESSURE DROP ACROSS  3 BEDS =  8.6 IN. H2O

TOTAL SYSTEM PRESSURE DROP = 14.8 IN. H2O

SPECIFIED   LIQUID-TO-GAS-RATIO    =  55. GAL/1000 ACF

LIMESTONE ADDITION = 0.3990E+05 LB/HR DRY LIMESTONE

CALCULATED LIMESTONE STOICHIOMETRY  =  1.24 MOLE CACO3 ADDED AS LIMESTONE
                                           PER MOLE SO2 ABSORBED

SOLUBLE CAO FROM FLY ASH =  0.0  MOLE PER MOLE SO2 ABSORBED

TOTAL SOLUBLE MGO        =  0.0  MOLE PER MOLE SO2 ABSORBED

TOTAL STOICHIOMETRY      =  1.24 MOLE SOLUBLE (CA+MG)
                                 PER MOLE SO2 ABSORBED

SCRUBBER INLET LIQUOR PH =  5.34

MAKE UP WATER =  680. GPM
```

```
SYSTEM SLUDGE DISCHARGE
------ ------ ---------
                                                        SOLID    LIQUID
                                                        COMP,    COMP,
SPECIES              LB-MOLE/HR     LB/HR               WT %     PPM

CASO3 .1/2 H2O       0.2182E+03     0.2817E+05          53.16
CASO4 .2H2O          0.9272E+02     0.1596E+05          30.11
CACO3                0.6629E+02     0.6635E+04          12.52
INSOLUBLES           ----------     0.2234E+04           4.22
H2O                  0.4350E+04     0.7837E+05
CA++                 0.9937E+01     0.3983E+03                    5010.
MG++                 0.0            0.0                              0.
SO3--                0.1254E+00     0.1004E+02                     126.
SO4--                0.8489E+00     0.8154E+02                    1026.
CL-                  0.1791E+02     0.6351E+03                    7988.

TOTAL DISCHARGE FLOW RATE = 0.1325E+06 LB/HR
                          =   200.     GPM

TOTAL DISSOLVED SOLIDS IN DISCHARGE LIQUID = 14150. PPM

DISCHARGE LIQUID PH =  7.24

SCRUBBER SLURRY BLEED
-------- ------- -----


TOTAL FLOW RATE = 0.3533E+06 LB/HR
                =    642.    GPM


TOTAL SUPERNATE RETURN
----- --------- ------


TOTAL FLOW RATE = 0.1820E+06 LB/HR
                =    364.    GPM
```

```
SUPERNATE TO WET BALL MILL
--------- -- --- ---- ----

TOTAL FLOW RATE = 0.2461E+05 LB/HR
                =     49.    GPM


LIMESTONE SLURRY FEED
--------- ------ ----

TOTAL FLOW RATE = 0.6651E+05 LB/HR
                =     84.    GPM

SUPERNATE RETURN TO SCRUBBER OR EHT
--------- ------ -- -------- -- ---

TOTAL FLOW RATE = 0.1574E+06 LB/HR
                =    315.    GPM

RECYCLE SLURRY TO SCRUBBER
------- ------ -- --------

TOTAL FLOW RATE = 0.3859E+08 LB/HR
                =  70087.    GPM

FLUE GAS COOLING SLURRY
---- --- ------- ------

TOTAL FLOW RATE = 0.2807E+07 LB/HR
                =   5097.    GPM
```

```
                   POND DFSIGN
                   ---- ------

  OPTIMIZED TO MINIMIZE TOTAL COST PLUS OVERHFAD


 POND DIMENSIONS
 ---- ----------

 DEPTH OF POND                  20.62 FT
 DEPTH OF EXCAVATION             3.24 FT
 LENGTH OF PERIMETER           13193.   FT
 LENGTH OF DIVIDER              2403.   FT

 AREA OF BOTTOM                 1074.   THOUSAND YD2
 AREA OF INSIDE WALLS            129.   THOUSAND YD2
 AREA OF OUTSIDE WALLS            96.   THOUSAND YD2
 AREA OF POND                   1194.   THOUSAND YD2
 AREA OF POND SITE              1343.   THOUSAND YD2
 AREA OF POND SITE               277.   ACRES

 VOLUME OF EXCAVATION           1241.   THOUSAND YD3
 VOLUME OF SLUDGE TO BE         7840.   THOUSAND YD3
 DISPOSED OVER LIFE OF PLANT    4860.   ACRE FT


 POND COSTS (THOUSANDS OF DOLLARS)
 ---- ----- ---------- -- --------

                                LABOR    MATERIAL    TOTAL
                                -----    --------    -----

 CLEARING LAND                   352.                 352.
 EXCAVATION                     1968.                1968.
 DIKE CONSTRUCTION               722.                 722.
 LINING( 10. IN. CLAY)           835.                 835.
 SODDING DIKE WALLS               48.        38.       86.
 ROAD CONSTRUCTION                 7.        14.       20.
 ---------------------------------------------------------
 POND CONSTRUCTION              3932.        52.     3984.
 LAND COST                                            971.
 ---------------------------------------------------------
 POND SITE                                           4955.
 OVERHEAD                                            2709.
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 TOTAL                                               7664.
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RAW MATERIAL HANDLING AND PREPARATION

INCLUDING 2 OPERATING AND 1 SPARE PREPARATION UNITS

ITEM	DESCRIPTION	NO.	MATERIAL	LABOR
CAR SHAKER AND HOIST	20HP SHAKER 7.5HP HOIST	1	28582.	1866.
CAR PULLER	25HP PULLER, 5HP RETURN	1	49345.	1866.
UNLOADING HOPPER	16FT DIA, 10FT STRAIGHT SIDE HT, CS	1	4180.	7711.
UNLOADING VIBRATING FEEDER	3.5HP	1	12134.	1866.
UNLOADING BELT CONVEYOR	20FT HORIZONTAL, 5HP	1	17527.	0.
UNLOADING INCLINE BELT CONVEYOR	310FT, 50HP	1	60670.	24875.
UNLOADING PIT DUST COLLECTOR	POLYPROPYLENE BAGTYPE, 2200 CFM,7.5HP	1	5258.	12438.
UNLOADING HOPPER	16FT DIA, 10FT STRAIGHT SIDE HT, CS	1	4180.	7711.
UNLOADING VIBRATING FEEDER	3.5HP	1	12134.	1866.
UNLOADING BELT CONVEYOR	20FT HORIZONTAL, 5HP	1	17527.	0.
UNLOADING INCLINE BELT CONVEYOR	310FT, 50HP	1	60670.	24875.
UNLOADING PIT DUST COLLECTOR	POLYPROPYLENE BAGTYPE, 2200 CFM,7.5HP	1	5258.	12438.
UNLOADING PIT SUMP PUMP	60GPM, 70FT HEAD, 5HP	1	3371.	746.
STORAGE BELT CONVEYOR	200FT, 5HP	1	57974.	16169.
STORAGE CONVEYOR TRIPPER	30FPM, 1HP	1	13482.	2488.
MOBILE EQUIPMENT	SCRAPPER TRACTOR	1.	136171.	0.
RECLAIM HOPPER	7FT WIDE, 4.25FT HT, 2FT WIDE BOTTOM, CS	2	1079.	1741.
RECLAIM VIBRATING FEEDER	3.5HP	2	24268.	3731.
RECLAIM BELT CONVEYOR	200FT, 5HP	1	40447.	8706.
RECLAIM INCLINE BELT CONVEYOR	193FT, 40HP	1	37750.	13930.
RECLAIM PIT DUST COLLECTOR	POLYPROPYLENE BAG TYPE	1	5258.	12438.
RECLAIM PIT SUMP PUMP	60GPM, 70FT HEAD, 5HP	1	3371.	746.
RECLAIM BUCKET ELEVATOR	90FT HIGH, 75HP	1	80894.	1617.
FEED BELT CONVEYOR	60.FT HORIZONTAL 7.5HP	1	20223.	1368.
FEED CONVEYOR TRIPPER	30 FPM, 1HP	1	13482.	2488.

FEED BIN	13FT DIA, 21FT STRAIGHT SIDE HT, COVERED, CS	3	16179.	29851.
BIN WEIGH FEEDER	14FT PULLEY CENTERS, 2HP	3	54603.	3731.
GYRATORY CRUSHERS	75HP	3	159765.	15672.
BALL MILL DUST COLLECTORS	POLYPROPYLENE BAG TYPE 2200 CFM, 7.5HP	3	15774.	37313.
BALL MILL	10.0TPH, 135.HP	3	442088.	40784.
MILLS PRODUCT TANK	5500 GAL 10FT DIA, 10FT HT, FLAKEGLASS LINED CS	3	14561.	22761.
MILLS PRODUCT TANK AGITATOR	10HP	3	24673.	1119.
MILLS PRODUCT TANK SLURRY PUMP	42.GPM, 60FT HEAD, 2.HP, 2 OPERATING AND 1 SPARES	3	7719.	1493.
SLURRY FEED TANK	44334.GAL, 19.6FT DIA, 19.6FT HT, FLAKEGLASS-LINED CS	1	10621.	22762.
SLURRY FEED TANK AGITATOR	41.HP	1	29197.	2155.
SLURRY FEED TANK PUMPS	21.GPM, 60 FT HEAD, 1.HP, 4 OPERATING AND 5 SPARE	9	22342.	4478.
TOTAL EQUIPMENT COST			1412984.	298907.

SCRUBBING

INCLUDING 4 OPERATING AND 2 SPARE SCRUBBING TRAINS

ITEM	DESCRIPTION	NO.	MATERIAL	LABOR
F.D. FANS	14.8IN H2O, WITH 1195. HP MOTOR AND DRIVE	6	2135508.	123875.
SHELL			974744.	
RUBBER LINING			1439962.	
MIST ELIMINATOR			442464.	
SLURRY HEADER AND NOZZLES			376418.	
GRIDS			566157.	
SPHERES			210822.	
TOTAL SCRUBBER COSTS		6	4010566.	334259.
REHEATERS		6	1256471.	51940.
SOOTBLOWERS		72	485362.	358206.
EFFLUENT HOLD TANK	231288.GAL, 34.0FT DIA, 34.0FT HT, FLAKEGLASS-LINED CS	6	109948.	206827.
EFFLUENT HOLD TANK AGITATOR	63.HP	6	415024.	183777.
COOLING SPRAY PUMPS	1274.GPM 100FT HEAD, 59.HP, 4 OPERATING AND 8 SPARE	12	141024.	20822.
ABSORBER RECYCLE PUMPS	8761.GPM, 100FT HEAD, 406.HP, 8 OPERATING AND 10 SPARE	18	790391.	62076.
MAKEUP WATER PUMPS	2549.GPM, 200.FT HEAD, 215.HP, 1 OPERATING AND 1 SPARE	2	19790.	1826.
			---------	---------
TOTAL EQUIPMENT COST			9364080.	1343607.

WASTE DISPOSAL

ITEM	DESCRIPTION	NO.	MATERIAL	LABOR
ABSORBER BLEED RECEIVING TANK	57761.GAL, 17.0FT DIA, 34.0FT HT, FLAKGLASS-LINED CS	1	14579.	31229.
ABSORBER BLEED TANK AGITATOR	36.HP	1	20467.	1511.
POND FEED SLURRY PUMPS	642.GPM, 130.FT HEAD 39.HP, 1 OPERATING AND 1 SPARE	2	14366.	2718.
POND SUPERNATE PUMPS	364.GPM, 192.FT HEAD, 29.HP, 1 OPERATING AND 1 SPARE	2	7144.	659.
TOTAL EQUIPMENT COST			56557.	36116.

LIMESTONE SLURRY PROCESS -- BASIS: 500 MW UNIT, 1978 STARTUP

PROJECTED CAPITAL INVESTMENT REQUIREMENTS - INDUSTRY BRIEFING-500 MW

	INVESTMENT, THOUSANDS OF 1977 DOLLARS				DISTRIBUTION
	RAW MATERIAL HANDLING AND PREPARATION	SCRUBBING	WASTE DISPOSAL	TOTAL	PERCENT OF DIRECT INVESTMENT
EQUIPMENT					
MATERIAL	1197.	7800.	46.	9042.	32.5
LABOR	255.	1116.	29.	1400.	5.0
PIPING					
MATERIAL	197.	2431.	729.	3357.	12.1
LABOR	79.	702.	270.	1052.	3.8
DUCTWORK					
MATERIAL	0.	1969.	0.	1969.	7.1
LABOR	0.	1347.	0.	1347.	4.8
FOUNDATIONS					
MATERIAL	104.	94.	10.	208.	0.7
LABOR	442.	281.	30.	754.	2.7
POND CONSTRUCTION	0.	0.	3984.	3984.	14.3
STRUCTURAL					
MATERIAL	227.	195.	1.	423.	1.5
LABOR	85.	436.	4.	525.	1.9
ELECTRICAL					
MATERIAL	141.	457.	83.	681.	2.4
LABOR	281.	701.	191.	1173.	4.2
INSTRUMENTATION					
MATERIAL	94.	743.	7.	844.	3.0
LABOR	21.	124.	2.	147.	0.5
BUILDINGS					
MATERIAL	28.	0.	0.	28.	0.1
LABOR	44.	0.	0.	44.	0.2
SERVICES AND MISCELLANEOUS	103.	595.	174.	872.	3.1
SUBTOTAL DIRECT INVESTMENT	3300.	18991.	5561.	27852.	100.0
ENGINEERING DESIGN AND SUPERVISION	297.	1709.	500.	2507.	9.0
CONSTRUCTION EXPENSES	528.	3039.	890.	4456.	16.0
CONTRACTOR FEES	165.	950.	278.	1393.	5.0
CONTINGENCY	330.	1899.	556.	2785.	10.0
SUBTOTAL FIXED INVESTMENT	4620.	26588.	7785.	38993.	140.0
ALLOWANCE FOR STARTUP AND MODIFICATIONS	370.	2127.	623.	3119.	11.2
INTEREST DURING CONSTRUCTION	554.	3191.	934.	4679.	16.8
SUBTOTAL CAPITAL INVESTMENT	5544.	31906.	9342.	46792.	168.0
LAND	7.	3.	979.	989.	3.6
WORKING CAPITAL	128.	738.	216.	1083.	3.9
TOTAL CAPITAL INVESTMENT	5680.	32647.	10538.	48864.	175.4

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LIMESTONE SLURRY PROCESS -- BASIS:  500 MW UNIT, 1978 STARTUP

PROJECTED REVENUE REQUIREMENTS - INDUSTRY BRIEFING-500 MW

                              DISPLAY SHEET FOR YEAR=     1
                             ANNUAL OPERATION KW-HR/KW =  7000

                        27.44 TONS PER HOUR                       DRY              SLUDGE
                            TOTAL FIXED INVESTMENT        48863000                    TOTAL
                                                                                      ANNUAL
    DIRECT_COSTS                                 ANNUAL_QUANTITY      UNIT_COST,$     COST,$

      RAW_MATERIAL

        LIMESTONE                                  146.3 K TONS        8.00/TON       1170200
        LIME                                         0.0 K TONS       50.00/TON      ________0

          SUBTOTAL RAW MATERIAL                                                       1170200

      CONVERSION_COSTS

        OPERATING LABOR AND
          SUPERVISION                            25990.0 MAN-HR      12.50/MAN-HR      324800
        UTILITIES
          STEAM                                 641550.0 K LB         2.00/K LB       1283100
          PROCESS WATER                         288140.0 K GAL        0.12/K GAL        34600
          ELECTRICITY                         48135440.0 KWH          0.029/KWH       1395900
        MAINTENANCE
          LABOR AND MATERIAL                                                          2029000
        ANALYSES                                  3760.0 HR          17.00/HR       _____63900

          SUBTOTAL CONVERSION COSTS                                                   5131300

          SUBTOTAL DIRECT COSTS                                                       6301500

    INDIRECT_COSTS

      DEPRECIATION                                                                    1559700
      COST OF CAPITAL AND TAXES, 17.20% OF UNDEPRECIATED INVESTMENT                   8404600
      INSURANCE & INTERIM REPLACEMENTS,  1.17% OF TOTAL CAPITAL INVESTMENT             571700
      OVERHEAD
        PLANT,  50.0% OF CONVERSION COSTS LESS UTILITIES                              1208900
        ADMINISTRATIVE, RESEARCH, AND SERVICE,
          10.0% OF OPERATING LABOR AND SUPERVISION                                  _____32500

          SUBTOTAL INDIRECT COSTS                                                    11777400

          SUBTOTAL ANNUAL REVENUE REQUIREMENT                                        18078900

          SLUDGE FIXATION COSTS              192100.0 TONS           12.00/TON       __2305200

          TOTAL ANNUAL REVENUE REQUIREMENT                                           _20384100

          EQUIVALENT UNIT REVENUE REQUIREMENT, MILLS/KWH                                  5.82
-----------------------------------------------------------------------------------------------------
    HEAT RATE    9000. BTU/KWH    -     HEAT VALUE OF COAL     10500 BTU/LB    -     COAL RATE   1500000 TONS/YR
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LIMESTONE SLURRY PROCESS -- BASIS: 500 MW UNIT, 1978 STARTUP

PROJECTED LIFETIME REVENUE REQUIREMENTS - INDUSTRY BRIEFING-500 MW

TOTAL CAPITAL INVESTMENT: $ 48864000

YEARS AFTER POWER UNIT START	ANNUAL OPERATION, KW-HR /KW	POWER UNIT HEAT REQUIREMENT, MILLION BTU /YEAR	POWER UNIT FUEL CONSUMPTION, TONS COAL /YEAR	SULFUR REMOVED BY POLLUTION CONTROL PROCESS, TONS/YEAR	BYPRODUCT RATE, EQUIVALENT TONS/YEAR DRY SLUDGE	SLUDGE FIXATION FEE $/TON DRY SLUDGE	ADJUSTED GROSS ANNUAL REVENUE REQUIREMENT EXCLUDING SLUDGE FIXATION COST, $/YEAR	TOTAL ANNUAL SLUDGE FIXATION COST, $/YEAR	NET ANNUAL INCREASE IN TOTAL REVENUE REQUIREMENT, $	CUMULATIVE NET INCREASE IN TOTAL REVENUE REQUIREMENT, $
1	7000	31500000	1500000	35000	192100	12.00	18078900	2305200	20384100	20384100
2	7000	31500000	1500000	35000	192100	12.00	17810600	2305200	20115800	40499900
3	7000	31500000	1500000	35000	192100	12.00	17542400	2305200	19847600	60347500
4	7000	31500000	1500000	35000	192100	12.00	17274100	2305200	19579300	79926800
5	7000	31500000	1500000	35000	192100	12.00	17005800	2305200	19311000	99237800
6	7000	31500000	1500000	35000	192100	12.00	16737500	2305200	19042700	118280500
7	7000	31500000	1500000	35000	192100	12.00	16469300	2305200	18774500	137055000
8	7000	31500000	1500000	35000	192100	12.00	16201000	2305200	18506200	155561200
9	7000	31500000	1500000	35000	192100	12.00	15932700	2305200	18237900	173799100
10	7000	31500000	1500000	35000	192100	12.00	15664500	2305200	17969700	191768800
11	5000	22500000	1071400	25000	137200	12.00	13622300	1646400	15268700	207037500
12	5000	22500000	1071400	25000	137200	12.00	13354000	1646400	15000400	222037900
13	5000	22500000	1071400	25000	137200	12.00	13085800	1646400	14732200	236770100
14	5000	22500000	1071400	25000	137200	12.00	12817500	1646400	14463900	251234000
15	5000	22500000	1071400	25000	137200	12.00	12549200	1646400	14195600	265429600
16	3500	15750000	750000	17500	96000	12.00	10877300	1152000	12029300	277458900
17	3500	15750000	750000	17500	96000	12.00	10609000	1152000	11761000	289219900
18	3500	15750000	750000	17500	96000	12.00	10340700	1152000	11492700	300712600
19	3500	15750000	750000	17500	96000	12.00	10072500	1152000	11224500	311937100
20	3500	15750000	750000	17500	96000	12.00	9804200	1152000	10956200	322893300
21	1500	6750000	321400	7500	41200	12.00	7471700	494400	7966100	330859400
22	1500	6750000	321400	7500	41200	12.00	7203500	494400	7697900	338557300
23	1500	6750000	321400	7500	41200	12.00	6935200	494400	7429600	345986900
24	1500	6750000	321400	7500	41200	12.00	6666900	494400	7161300	353148200
25	1500	6750000	321400	7500	41200	12.00	6398600	494400	6893000	360041200
26	1500	6750000	321400	7500	41200	12.00	6130400	494400	6624800	366666000
27	1500	6750000	321400	7500	41200	12.00	5862100	494400	6356500	373022500
28	1500	6750000	321400	7500	41200	12.00	5593800	494400	6088200	379110700
29	1500	6750000	321400	7500	41200	12.00	5325600	494400	5820000	384930700
30	1500	6750000	321400	7500	41200	12.00	5057300	494400	5551700	390482400
OT	127500	573750000	27321000	637500	3499000		348494400	41988000	390482400	

LIFETIME AVERAGE INCREASE IN UNIT REVENUE REQUIREMENT			
DOLLARS PER TON OF COAL BURNED	12.76	1.53	14.29
MILLS PER KILOWATT-HOUR	5.47	0.66	6.13
CENTS PER MILLION BTU HEAT INPUT	60.74	7.32	68.06
DOLLARS PER TON OF SULFUR REMOVED	546.66	65.86	612.52
EVENUE REQUIREMENT DISCOUNTED AT 11.6% TO INITIAL YEAR, DOLLARS	125705100	16366000	142071100
LEVELIZED INCREASE IN UNIT REVENUE REQUIREMENT EQUIVALENT TO DISCOUNTED REQUIREMENT OVER LIFE OF POWER UNIT			
DOLLARS PER TON OF COAL BURNED	11.80	1.54	13.34
MILLS PER KILOWATT-HOUR	5.06	0.66	5.72
CENTS PER MILLION BTU HEAT INPUT	56.21	7.32	63.53
DOLLARS PER TON OF SULFUR REMOVED	505.86	65.86	571.71

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LIME PREPARATION AREA
EQUIPMENT LIST

INCLUDING 1 OPERATING AND 1 SPARE PREPARATION UNITS

ITEM	DESCRIPTION	NO.	MATERIAL	LABOR
CONVEYOR FROM CALCINATION PLANT	1500FT HORIZONTAL, 30HP	1	153293.	46144.
STORAGE SILO ELEVATOR	29.FT HIGH, 50 HP	1	32047.	650.
CONCRETE STORAGE SILO	659.FT3, 8.2FT DIA , 12.4FT STRAIGHT SIDE STORAGE HT	1	3161.	7624.
STORAGE SILO HOPPER BOTTOM	60 DEGREE, CS	1	401.	1194.
RECLAIM VIBRATING FEEDER	3.5HP	1	12134.	1866.
RECLAIM BELT CONVEYOR	83.FT HORIZONTAL, 5HP	1	15232.	3953.
FEED BIN ELEVATOR	50FT HIGH, 50HP	1	52311.	995.
FEED BIN	10FT DIA, 15FT STRAIGHT SIDE HT, COVERED, CS	2	5393.	9950.
BIN.VIBRATING FEEDER	3.5HP	2	9168.	3234.
BIN WEIGH FEEDER	12FT, 12IN SCREW, 1HP	2	11864.	1244.
SLAKER	0.TPH, 0.HP	2	24960.	2303.
SLAKER PRODUCT TANK		2	12134.	18905.
SLAKER PRODUCT TANK AGITATOR	10HP	2	14291.	746.

LIME PREPARATION AREA
EQUIPMENT LIST (CONTINUED)

LIME SYSTEM DUST COLLECTORS	POLYPROPYLENE BAG TYPE 2200 CFM,7.5HP	4	21032.	49751.
SLAKER PRODUCT TANK SLURRY PUMPS	1.GPM, 60FT HEAD, 0.HP, 1 OPERATING AND 1 SPARES	2	4448.	995.
SLURRY FEED TANK	680.GAL, 4.9FT DIA, 4.9FT HT, FLAKEGLASS-LINED CS	1	656.	1406.
SLURRY FEED TANK AGITATOR	1.HP	1	2867.	212.
SLURRY FEED TANK PUMPS	0.GPM, 60 FT HEAD, 0.HP, 4 OPERATING AND 5 SPARE	9	19444.	4478.
			---------	---------
TOTAL EQUIPMENT COST			394838.	155649.

SCRUBBING

INCLUDING 4 OPERATING AND 2 SPARE SCRUBBING TRAINS

ITEM	DESCRIPTION	NO.	MATERIAL	LABOR

WASTE DISPOSAL AREA
OPTIONS 2 AND 3
EQUIPMENT LIST

WASTE DISPOSAL

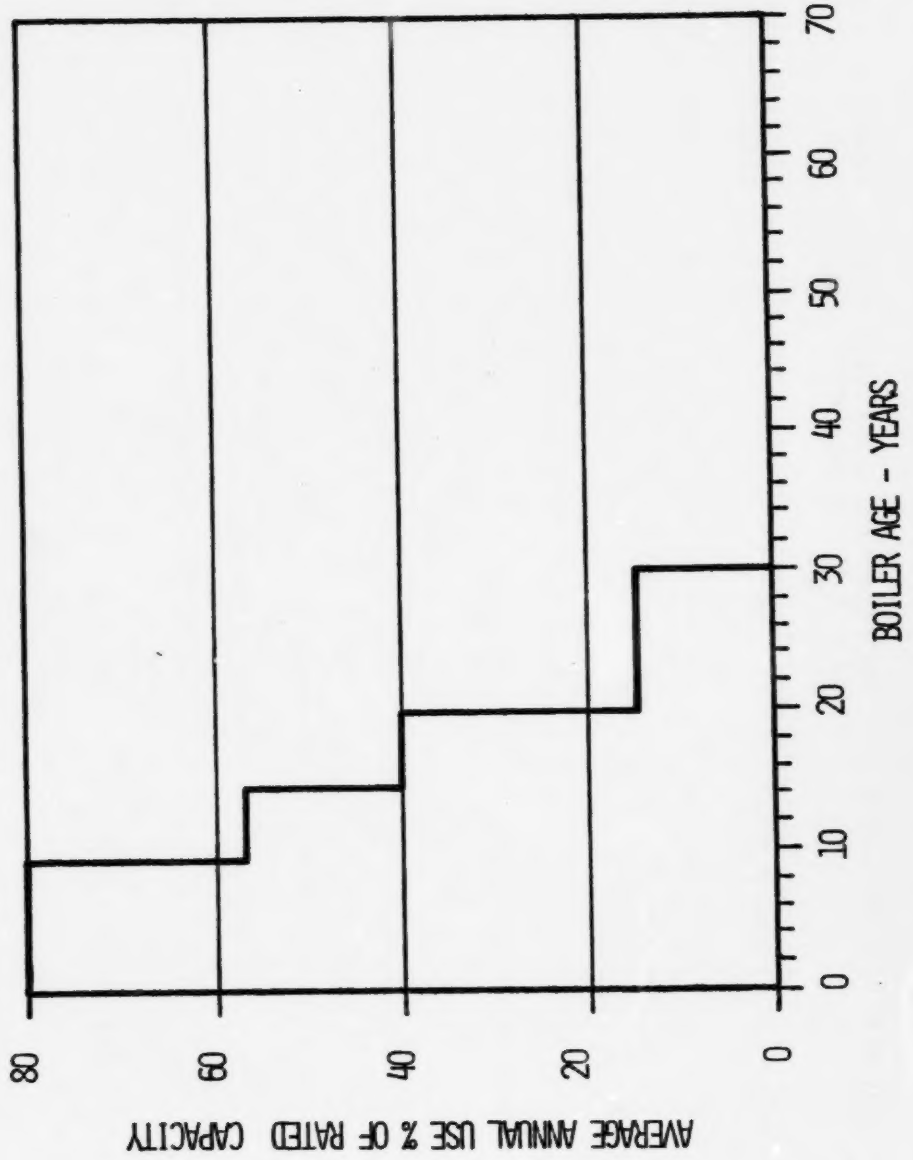
ITEM	DESCRIPTION	NO.	MATERIAL	LABOR
ABSORBER BLEED RECEIVING TANK	52510.GAL., 16.5FT DIA, 32.9FT HT, FLAKGLASS-LINED CS	1	13682.	29306.
ABSORBER BLEED TANK AGITATOR	34.HP	1	18967.	1400.
THICKENER FEED PUMP	665.GPM, 60FT HEAD, 18.HP, 1 OPERATING AND 1 SPARE	2	13078.	2752.
THICKENER	18248.SQ.FT.,152.FT DIA, 8.6FT HT	1	536285.	366827.
THICKENER OVERFLOW PUMPS	438.GPM, 75.0FT HEAD, 14.HP, 1 OPERATING AND 1 SPARE	2	5913.	545.
THICKENER OVERFLOW TANK	7224.GAL., 12.0FT DIA, 8.6FT HT	1	1637.	3927.
SLUDGE FIXATION FEED PUMP	207.GPM, 50FT HEAD, 6.HP, 1 OPERATING AND 1 SPARE	2	10575.	1816.
TOTAL EQUIPMENT COST			600137.	406574.

WASTE DISPOSAL AREA
OPTION 4
EQUIPMENT LIST

WASTE DISPOSAL

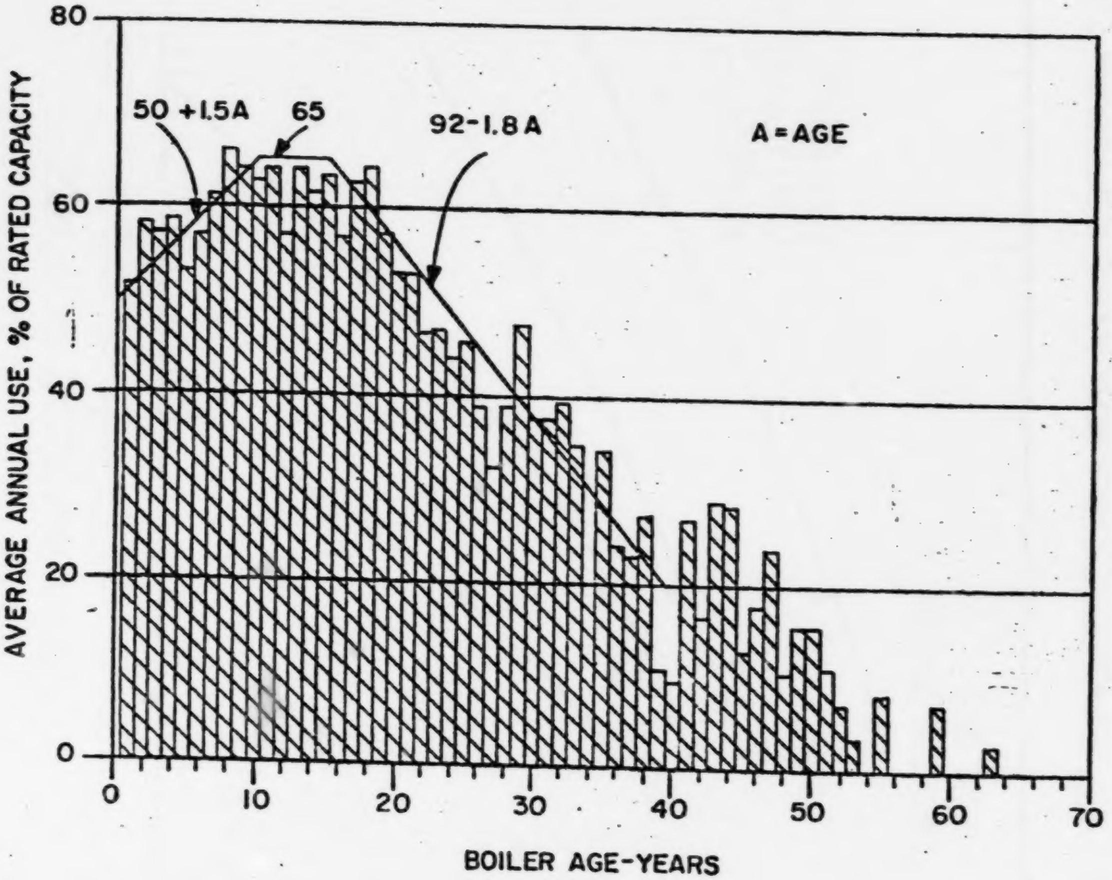
ITEM	DESCRIPTION	NO.	MATERIAL	LABOR
ABSORBER BLEED RECEIVING TANK	52510.GAL, 16.5FT DIA, 32.9FT HT, FLAKGLASS-LINED CS	1	13682.	29306.
ABSORBER BLEED TANK AGITATOR	34.HP	1	18967.	1400.
THICKENER FEED PUMP	652.GPM, 60FT HEAD, 18.HP, 1 OPERATING AND 1 SPARE	2	13053.	2734.
THICKENER	17915.SQ.FT.,151.FT DIA, 8.5FT HT	1	532294.	362329.
THICKENER OVERFLOW PUMPS	430.GPM, 75.0FT HEAD, 14.HP, 1 OPERATING AND 1 SPARE	2	5877.	542.
THICKENER OVERFLOW TANK	7092.GAL, 11.9FT DIA, 8.5FT HT	1	1618.	3880.
FILTER FEED SLURRY PUMP	102.GPM, 50FT HEAD, 3.HP, 2 OPERATING AND 1 SPARE	3	10750.	2114.
FILTER	269.SQ FT FILTRATION AREA	2	180058.	16611.
FILTRATE PUMP (PER FILTER)	45.GPM, 20.0FT HEAD, 0.HP, 2 OPERATING AND 2 SPARE	4	6211.	573.
FILTRATE SURGE TANK	1482.GAL, 6.3FT DIA, 6.3FT HT	1	594.	1424.
FILTRATE SURGE TANK PUMP	90.GPM, 85.0FT HEAD, 3.HP, 1 OPERATING AND 1SPARE	2	3874.	357.

FIVE PROCESS STUDY OPERATING PROFILE

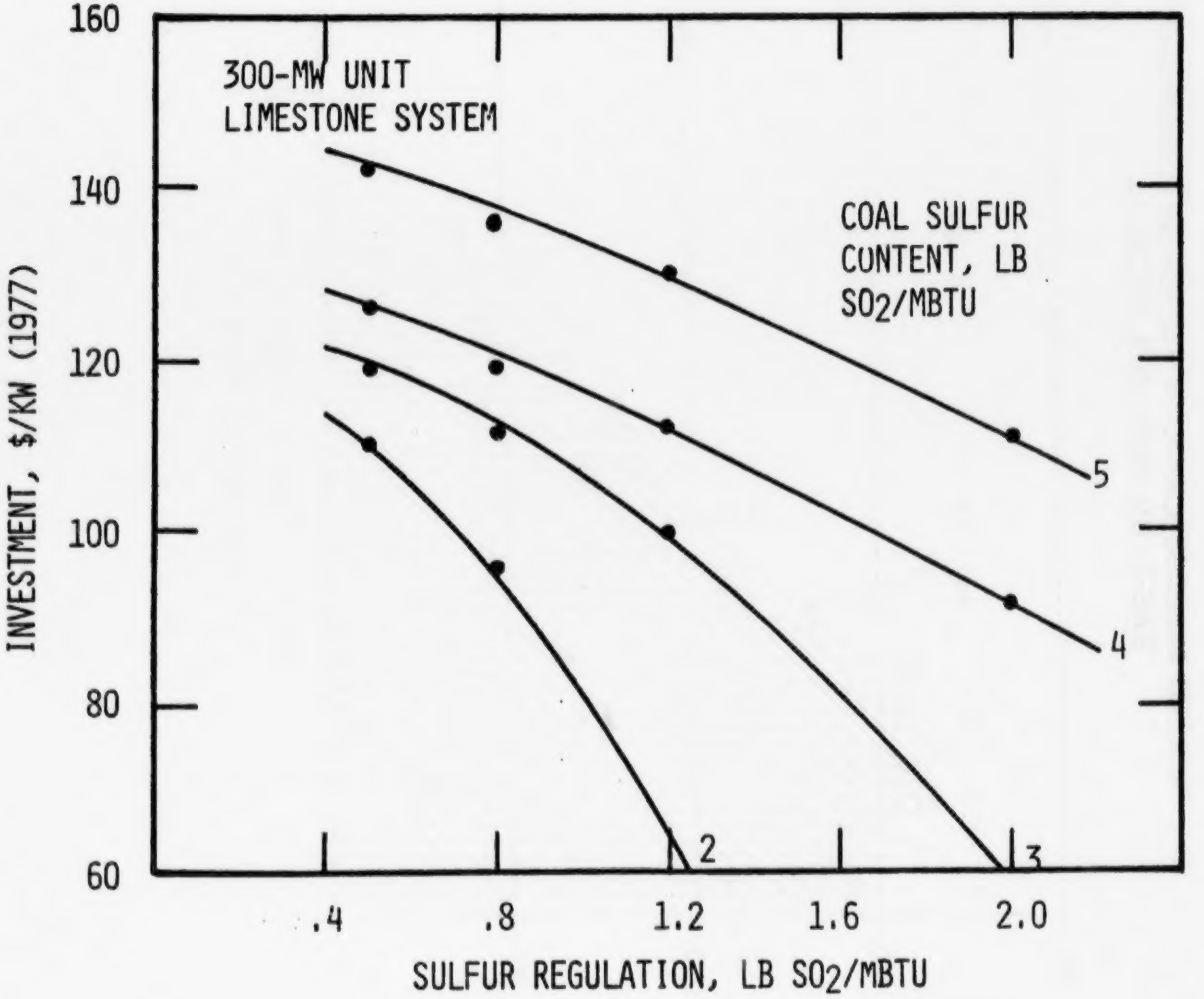


FPC OPERATING PROFILE

ALL BOILERS AVERAGE CAPACITY FACTOR vs. BOILER AGE-
BASED ON 1969-1973 FPC DATA



NATIONAL ECONOMIC RESEARCH ASSOCIATE'S SENSITIVITY STUDY



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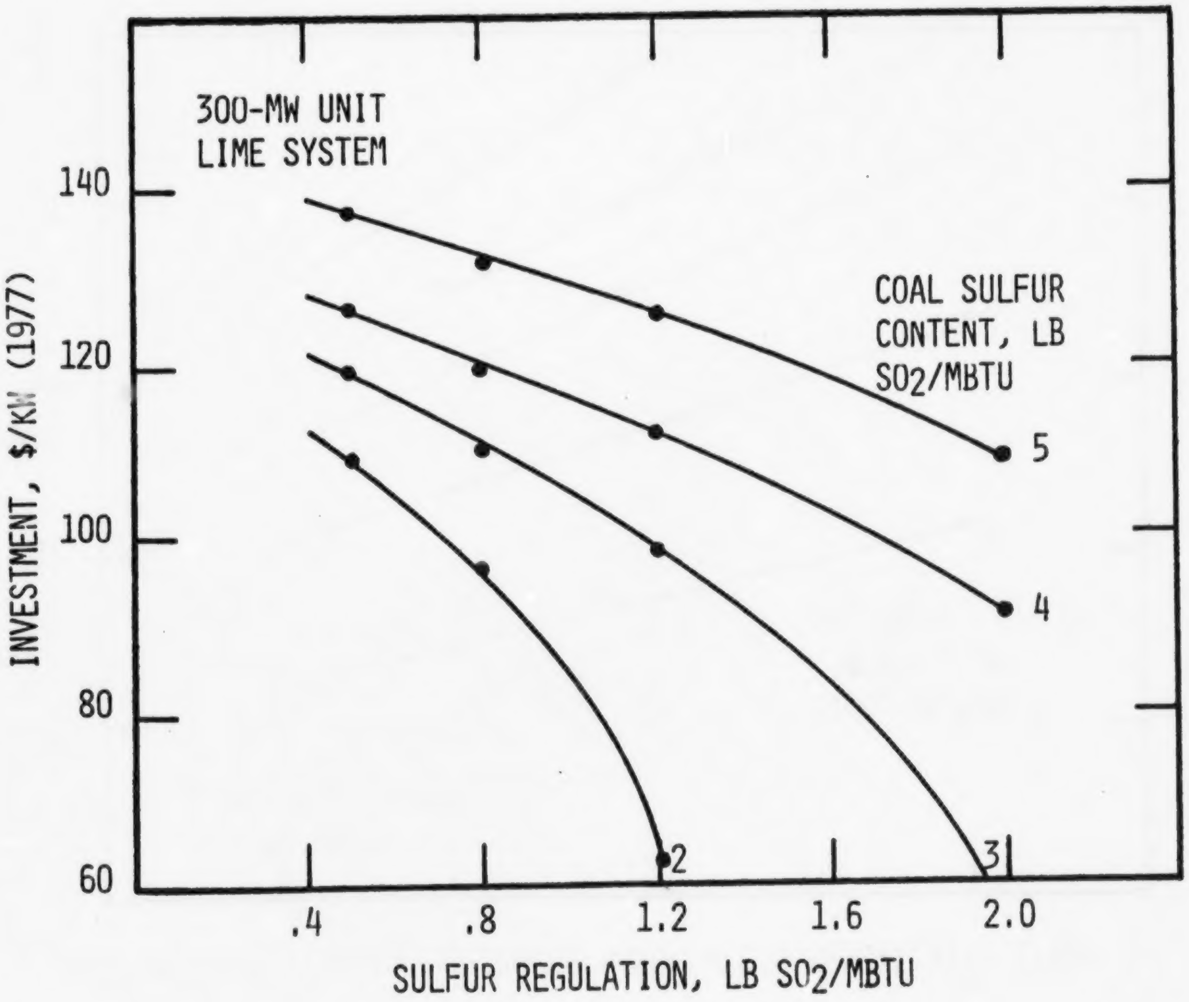
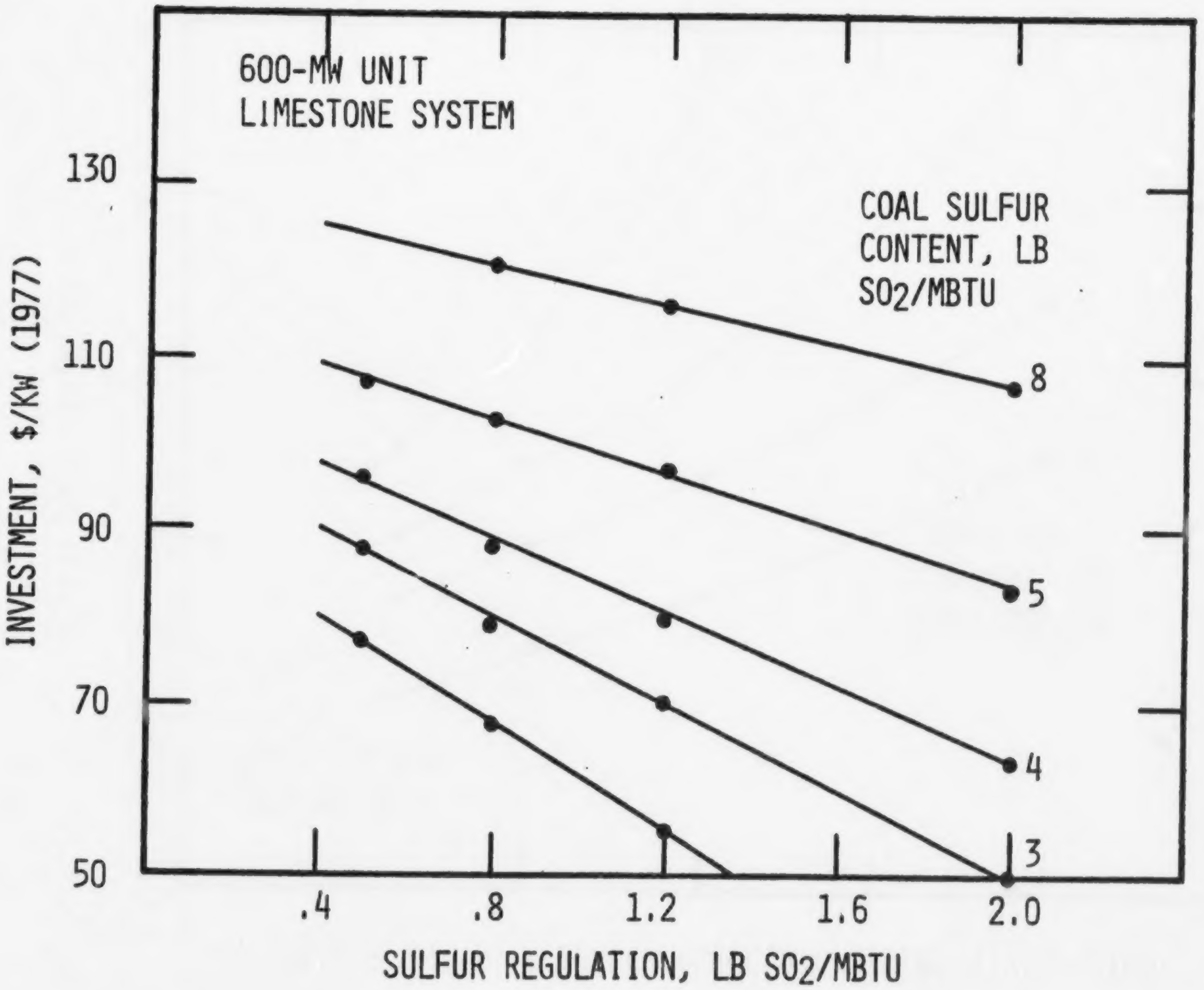
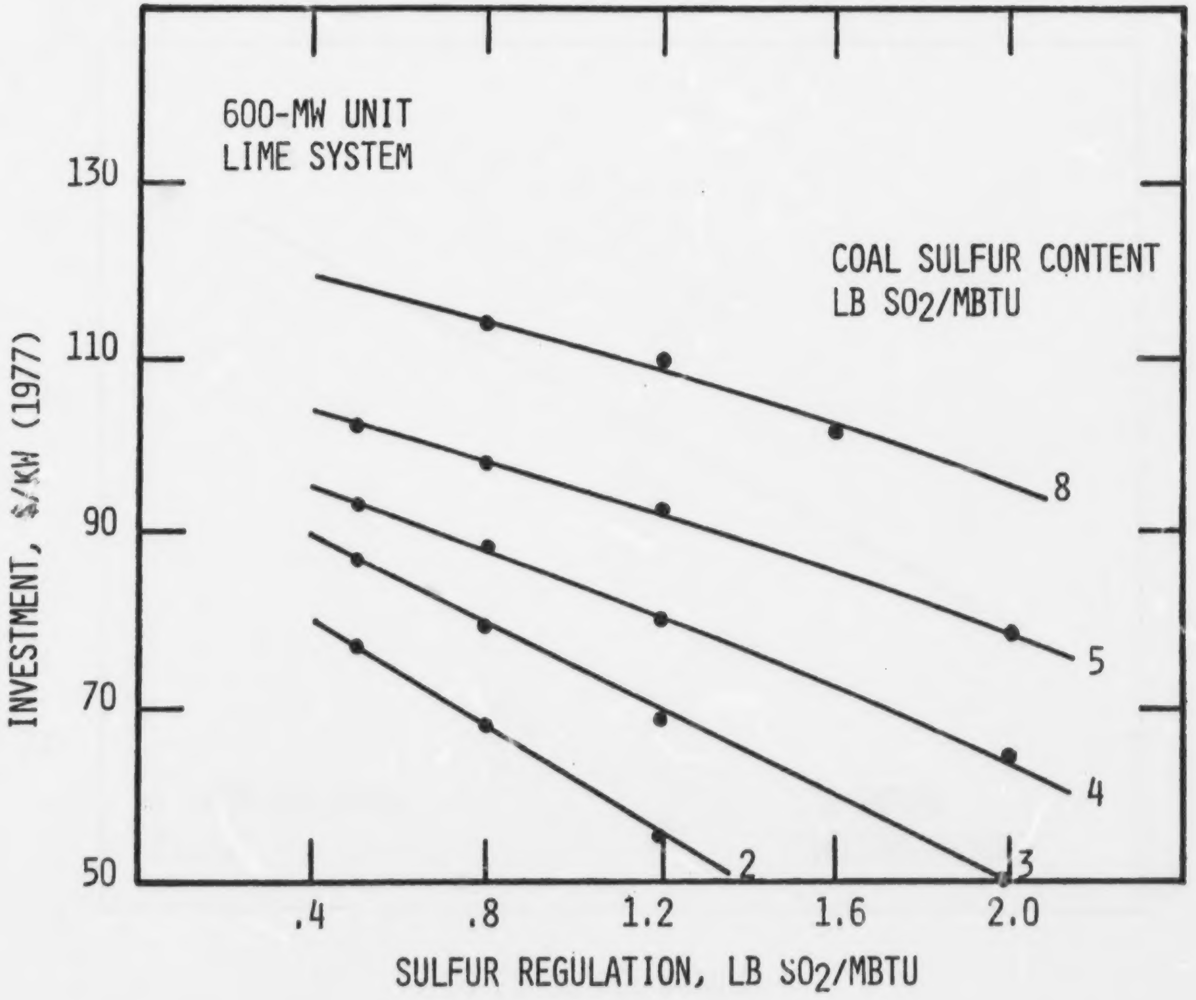
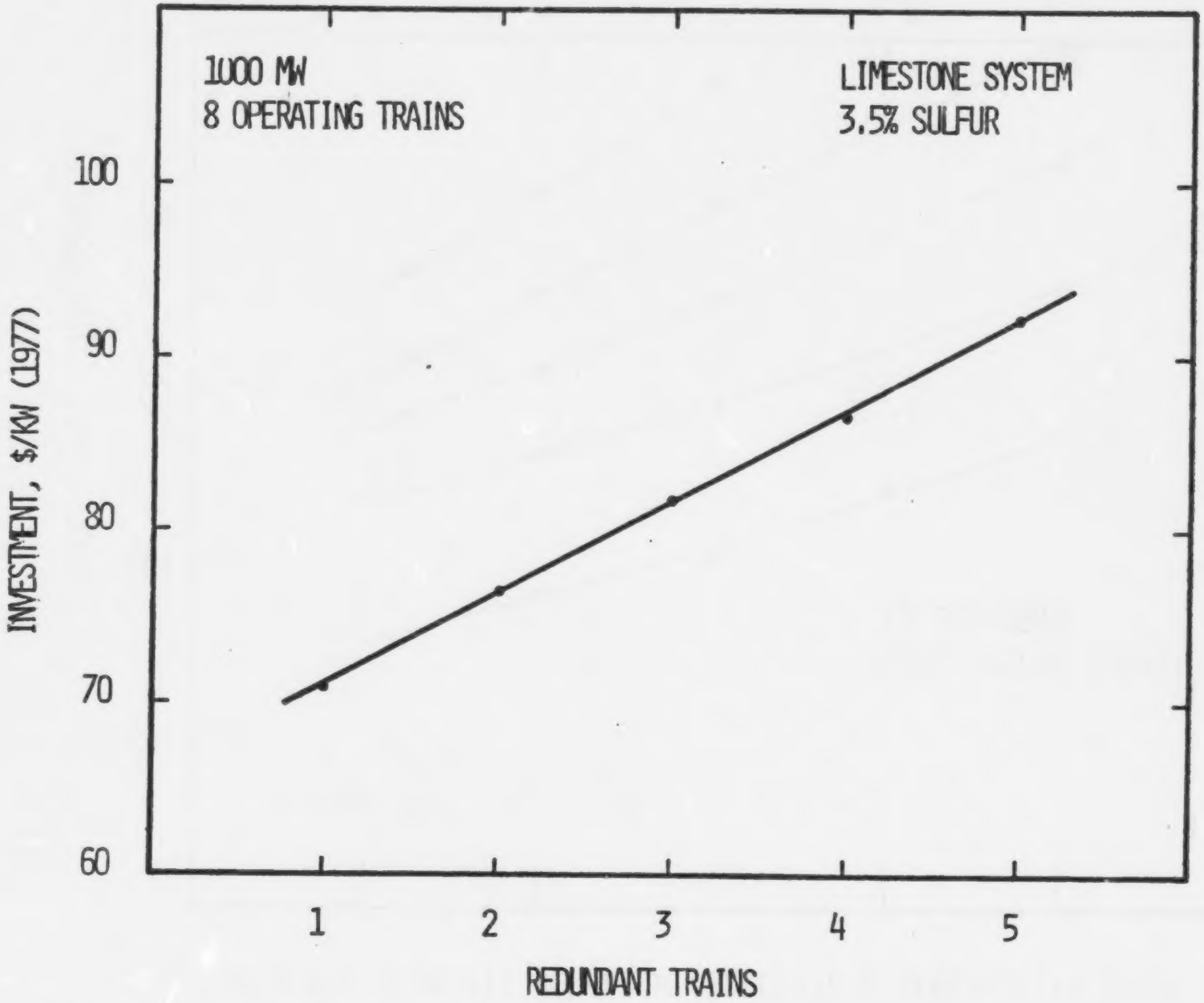


NATIONAL ECONOMIC RESEARCH ASSOCIATE'S SENSITIVITY STUDY

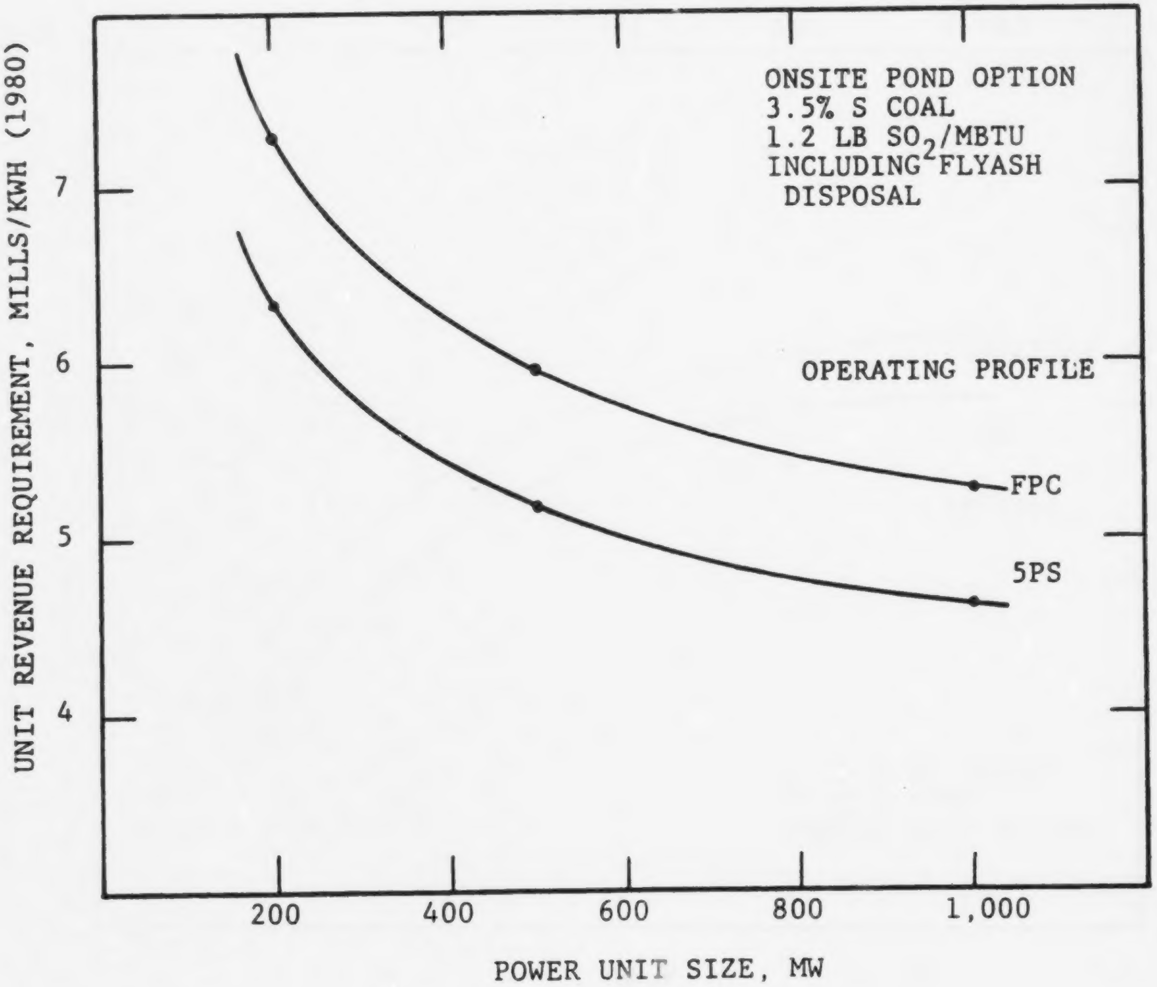


NATIONAL ECONOMIC RESEARCH ASSOCIATE'S SENSITIVITY STUDY

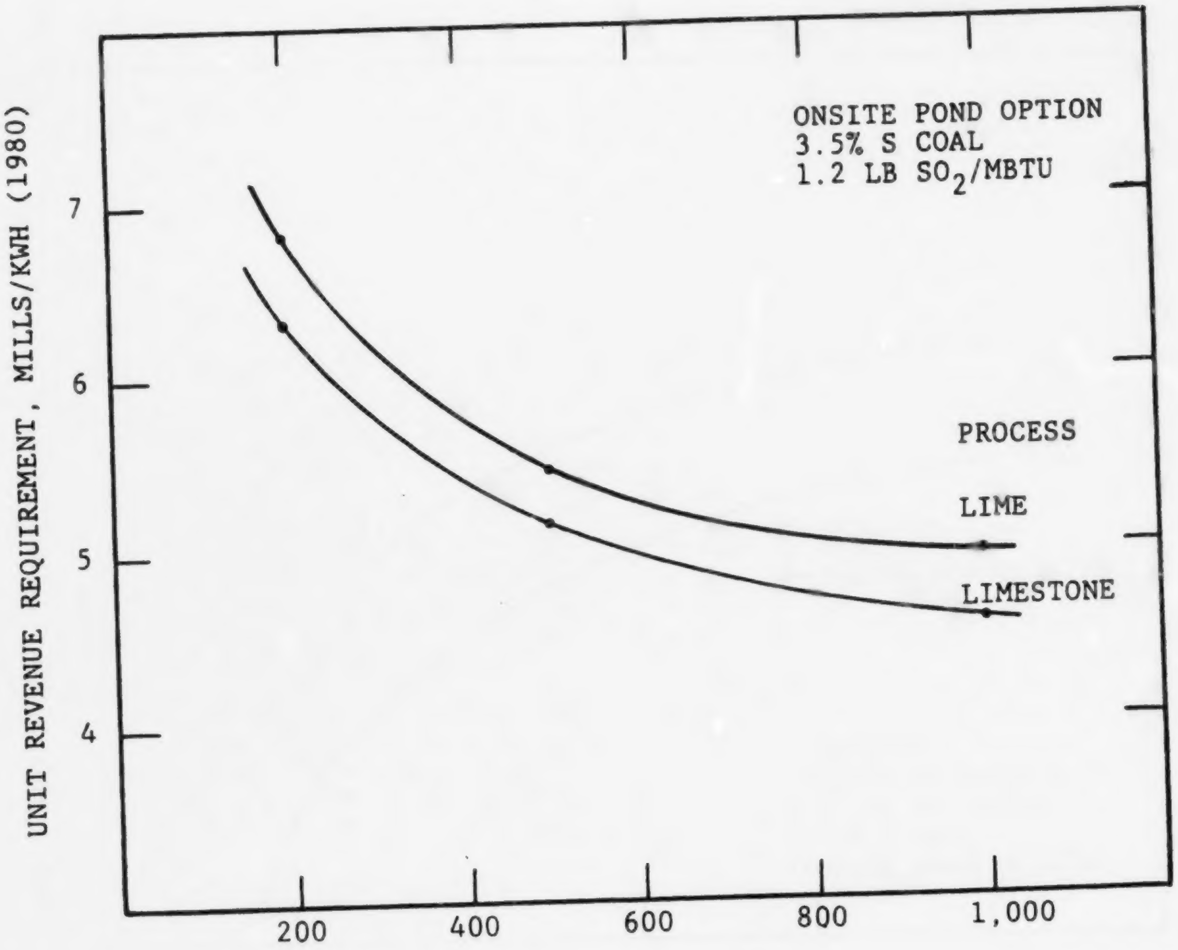




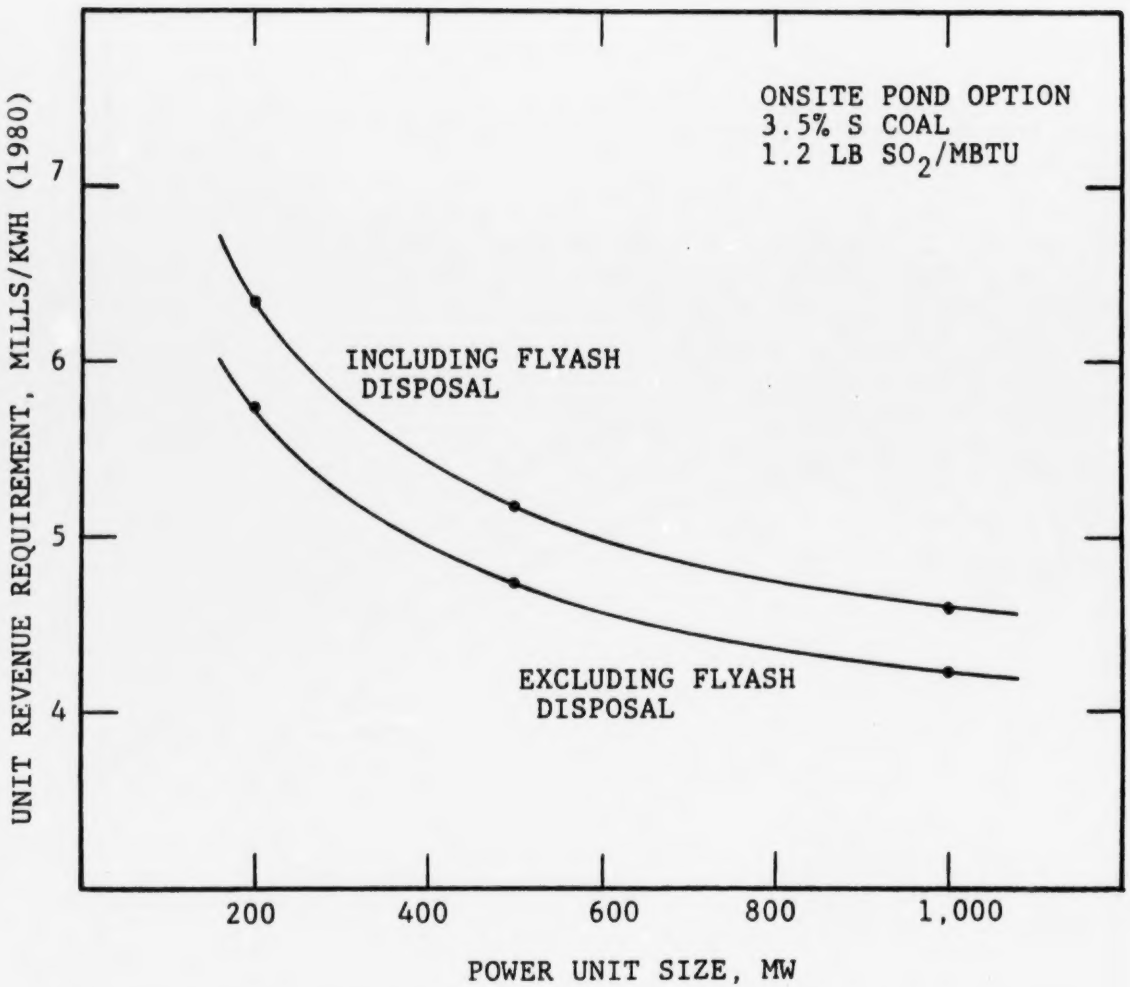
COSTS OF REDUNDANCY



LIMESTONE PROCESS - EFFECT OF POWER UNIT SIZE AND OPERATING PROFILE ON UNIT REVENUE REQUIREMENT



EFFECT OF POWER UNIT SIZE AND PROCESS
ON UNIT REVENUE REQUIREMENT



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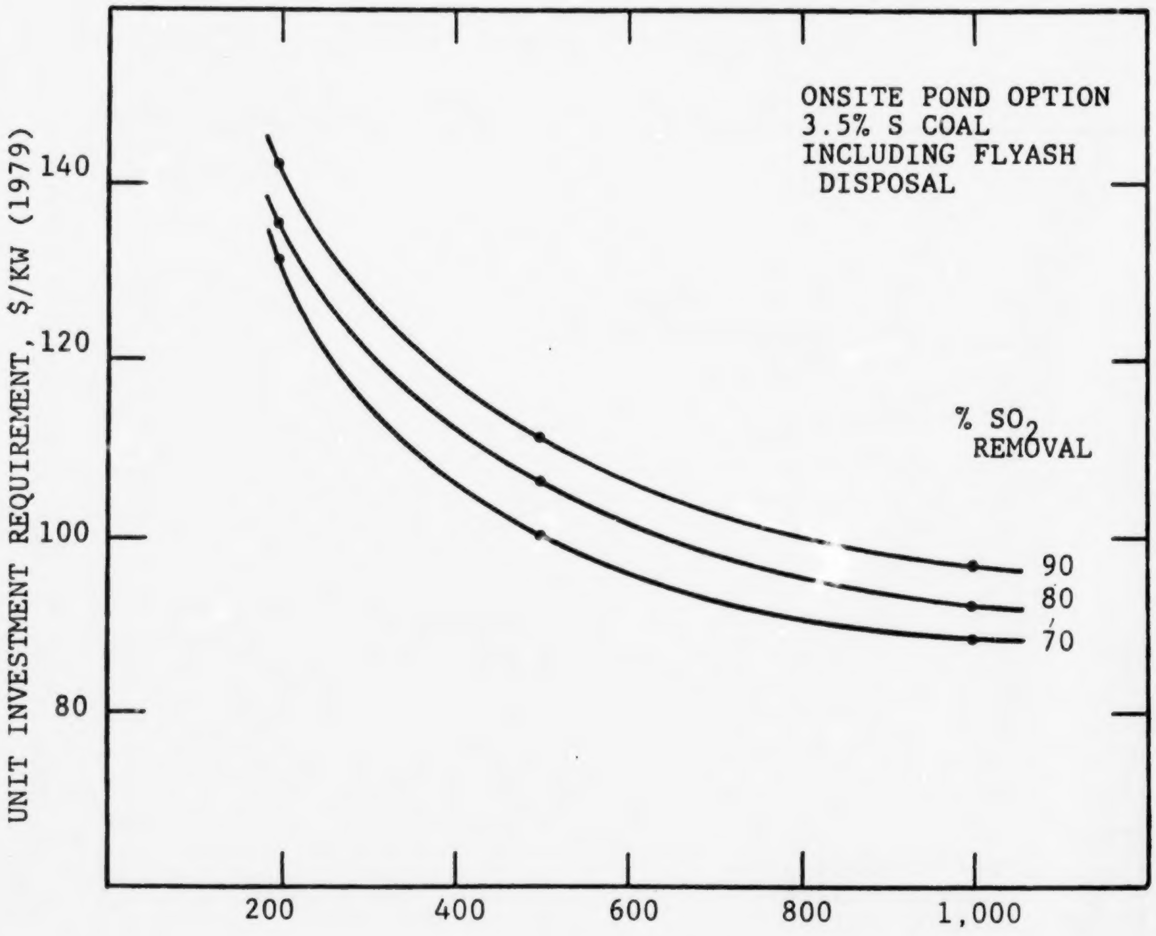
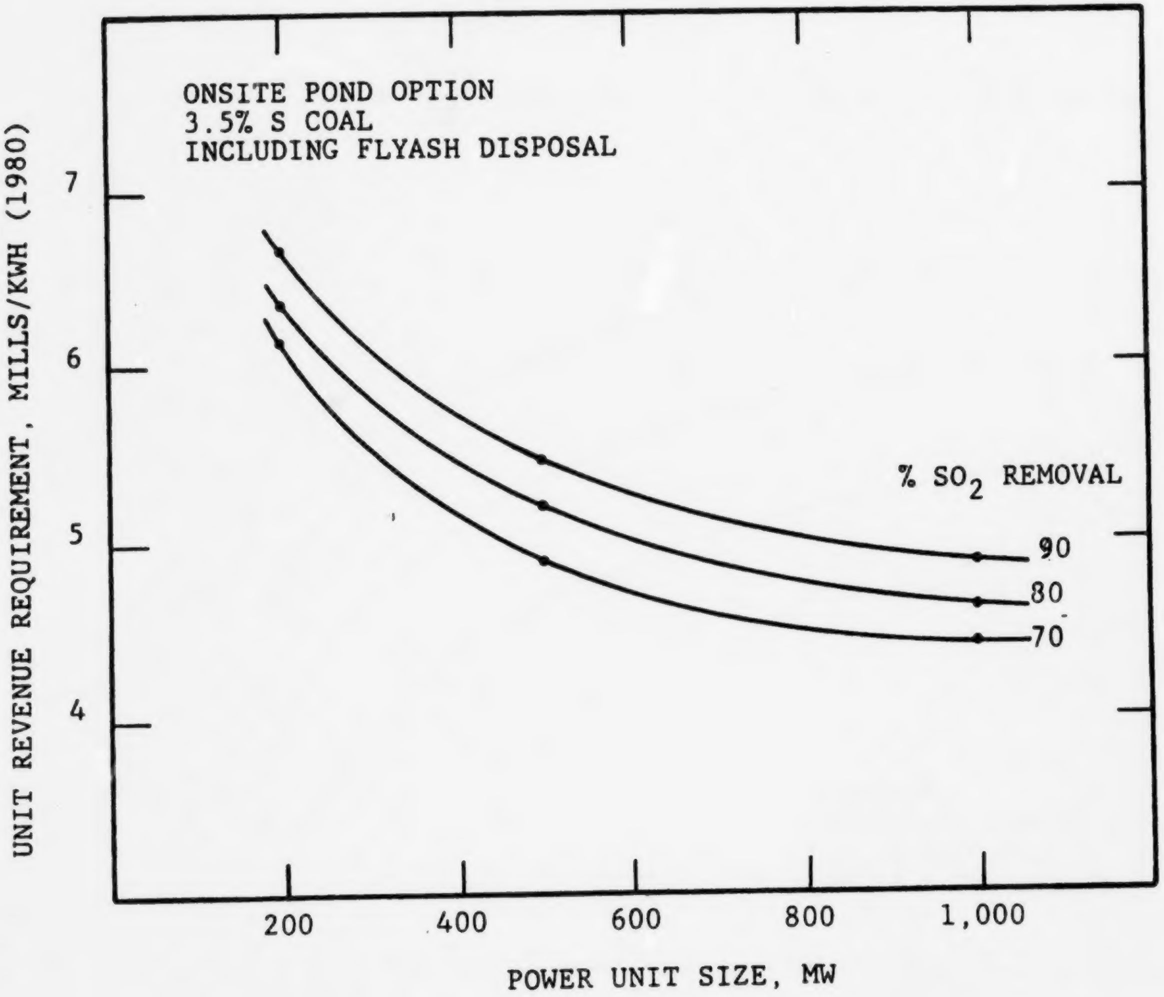
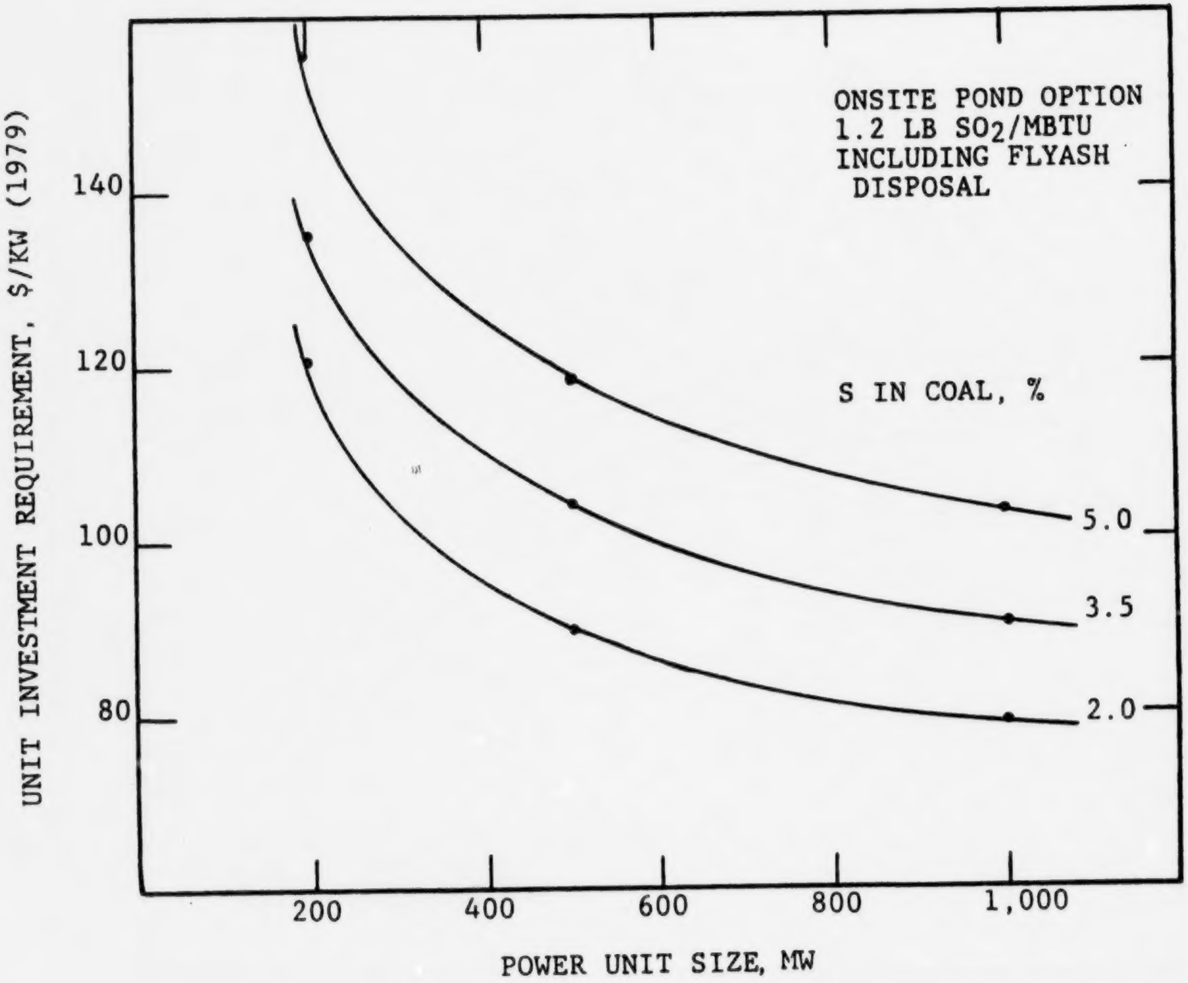


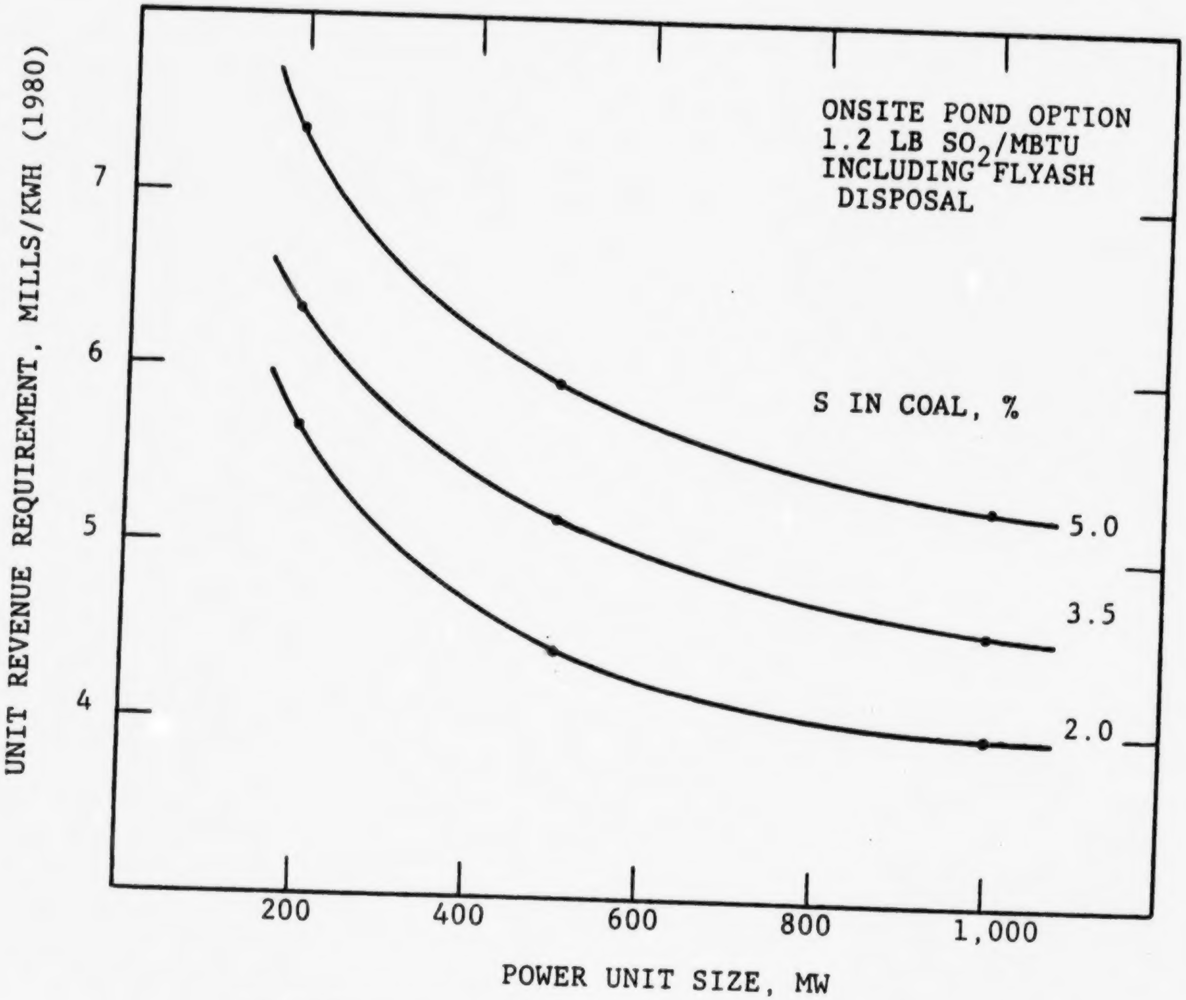
LIMESTONE PROCESS - EFFECT OF POWER UNIT SIZE AND % SO_2 REMOVAL ON UNIT INVESTMENT REQUIREMENT



LIMESTONE PROCESS - EFFECT OF POWER UNIT SIZE AND % SO_2 REMOVAL ON UNIT REVENUE REQUIREMENT



LIMESTONE PROCESS - EFFECT OF POWER UNIT SIZE AND S IN COAL ON UNIT INVESTMENT REQUIREMENT



LIMESTONE PROCESS - EFFECT OF POWER UNIT SIZE AND S IN COAL ON UNIT REVENUE REQUIREMENT

COAL CLEANING VERSUS 100% FGD

CASE 1 ROM COAL → 2000 MW → FGD
CASE 2 ROM COAL → PREP PLT → 2000 MW → FGD
CASE 3 ROM COAL → CHEM COMM & PREP PLT → 2000 MW → FGD

	CASE 1	CASE 2	CASE 3
CAPACITY FACTOR	45	56	56
GENERATION (KWH/YR)	7.88×10^9	9.81×10^9	9.81×10^9
PRODUCTION COSTS (MILLS/KWH)	15.2	12.7	12.9
FGD REVENUE REQUIREMENT (MILLS/KWH)	7.1	6.1	5.8
TOTAL GENERATION COSTS (MILLS/KWH)	22.3	18.8	18.7

LANDFILL AND PONDING CONCEPTS FOR FGD SLUDGE DISPOSAL

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ABSTRACT

This paper is concerned with the environmentally sound disposal of flue gas desulfurization (FGD) sludges. The environmental considerations and the technology and costs associated with the disposal of FGD sludges by landfilling and ponding are summarized. Concepts discussed are lined ponds, unlined ponds equipped with underdrainage, chemical treatment and landfilling, and conversion to gypsum. The need for environmental control is reviewed. The capabilities of each concept to prevent water pollution and the environmental considerations that require site maintenance are discussed. Bearing strengths associated with landfill concepts are included, and the status of developments of non-operational concepts, i.e., ponding with underdrainage and the disposal of FGD gypsum, are discussed. Additionally, disposal site volume requirements and estimated disposal costs are given.

Introduction

This paper summarizes current concepts of landfilling and ponding for the environmentally sound disposal of flue gas desulfurization (FGD) sludges. The techniques discussed herein represent the results of studies and assessments performed by The Aerospace Corporation under contract to the Industrial Environmental Research Laboratory of the U.S. Environmental Protection Agency (EPA), Research Triangle Park, North Carolina. These techniques do not constitute endorsement or approval by the EPA, but are presented as the authors' assessments of the best available methods for the disposal of FGD sludges by landfilling or ponding.

With the passage of the Resource Conservation and Recovery Act (RCRA), public law 94-580, October 1976, guidelines and criteria are forthcoming for application to FGD sludges. Determinations will be made by the EPA as to whether these sludges are to be considered hazardous, and, depending on those determinations, criteria will be developed for FGD sludge disposal.

Without federal criteria applicable specifically to FGD sludges, almost all studies and developments up to this time have used drinking water criteria as the basis for establishing requirements for disposal. Because the trace element and salt content of most samples analyzed exceeded the drinking water criteria at least for some of the constituents, the general approach taken has been to dispose of FGD sludges such that no direct discharge to any water supply would be permitted, that any seepage would be minimized or perhaps totally eliminated, and that runoff would be controlled. Additionally, a strong effort has been made by industrial and government agency development and evaluation programs to determine disposal techniques that would not only be environmentally sound from the standpoint of water quality control and, when practical, would also result in reclamation of the land area selected for the disposal site. As a result, all disposal techniques that have been developed for FGD sludge are intended for the control of water quality, but not all of them produce reclaimable disposal sites. The techniques discussed herein consist of the following: (a) ponding of untreated sludges, (b) disposal of untreated sludges in ponds equipped with underdrainage systems, (c) chemical treatment and landfilling, and (d) conversion to gypsum and subsequent disposal.

The basic characteristics of each of these approaches are discussed as to protection of water supplies, land reclamation, and disposal costs.

Water Quality Criteria

A comparison of chemical constituents from a large number of analyses of sludge liquors in a discharge stream with the National Interim Primary Drinking Water Regulations (40 CFR 141) is given in Table I as a ratio of constituent concentration to water criteria. These ratios are given for the range of constituents from a composite of data for ten eastern and western sludges, with and without fly ash, and for the ten independent samples. It should be noted that the values used in this comparative analysis represent the initial concentrations that would seep from the base of an untreated sludge pile.

Table I. Comparison of Sludge Liquors with Water Criteria

NIPDWR Drinking Water Criteria, mg/ℓ		Concentration ÷ Criteria (Nondimensional)										
		Range of All Samples	Sample[a]									
			A	B	C	D	E	F	G	H	I	J
As	0.05	< 0.8 - 2.8	0.6	0.4	2.0	0.04	0.4	1.2	2.8	0.1	0.8	0.2
Cd	0.01	0.4 - 11	5.0	1.2	0.4	--	11	1.3	--	--	5	2.5
Cr	0.05	0.22 - 5	5.0	0.8	1.8	--	0.6	0.2	--	--	--	1.1
Pb	0.05	0.2 - 6.6	0.8	3.0	4.6	<0.2	6.6	0.2	< 0.2	< 0.2	0.8	< 0.1
Hg	0.002	0.03 - 2.5	2.5	--	--	<0.1	< 0.5	< 0.001	< 0.1	< 0.1	0.1	0.03
Se	0.01	0.28 - 20	10.0	3.3	10.0	4.2	< 2	7.8	20	14	2.8	0.3
F	2	< 0.5 - 5	--	0.5	3.3	--	1.7	1	--	--	5	< 0.5
TDS	500	6.6 - 48.5	36	6.6	30.0	13.4	18.8	20.5	28	18.4	8.4	48.5
pH (actual values)[b]		6.7 - 12.2	6.7	6.8	8.0	12.2	8.7	8.0	7.8	7.3	10.7	8.9

[a] Sample data are as follows:

Sample	Station	Absorbent	% Ash	Sampling Date
A	Mohave	Limestone	3	Mar 1973
B	Cholla	Limestone	59	Nov 1974
C	Shawnee	Limestone	40	Jun 1974
D	Shawnee	Limestone	6	Jan 1977
E	Shawnee	Lime	40	Jun 1974
F	Shawnee	Lime	6	Sep 1976
G	Shawnee	Lime	6	Oct 1976
H	Shawnee	Limestone[c]	6	Aug 1977
I	Duquesne Phillips	Lime	60	Jun 1974
J	LG& E Paddy's Run	Carbide lime	12	Jul 1976

[b] EPA-proposed secondary regulation is 6.5 to 8.5.

[c] Forced-oxidized to gypsum.

In Table I, the ratios for the range of constituent concentrations of the composite data show that all elements analyzed, as well as the total dissolved solids (TDS) and pH, exceed drinking water criteria. However, in observing the ratios for the ten independent samples shown in the table, it can be seen that, except for selenium in two samples and cadmium in one, no trace element exceeds the criteria by a factor greater than 10. (Water criteria for barium, nitrate, and silver are 1, 10, and 0.05 mg/ℓ, respectively. Limited field evaluation leachate data show maximum concentrations for these elements to be about 5, 1, and 0.5 times the criteria, respectively.) The TDS are high for most of the samples, and the pH is excessive for two of the samples. Although trace elements are not eliminated as a matter of concern for some sites by these data, there are indications that in many cases the concentrations are quite low and that, generally, the concern may be for the concentration of dissolved solids and, in some cases, pH. Chemical oxygen demand (COD) was considered somewhat differently. Values of COD in fresh sludge ranged between 40 and 140 mg/ℓ, but, because of the rapid oxidation characteristic of sulfite sludge, the COD after one pore volume displacement by leaching was 10 mg/ℓ or less and rapidly decreasing. Therefore, because COD is significant only for fresh sludge and because the sludge is not discharged directly to streams, it was concluded that COD is not a critical parameter.

Because of depletion of the material with leaching time, cation exchange and adsorption in the soil, and dilution between the disposal site and the consumer tap, it is difficult at this time to specifically quantify the degree of pollution potential at a given site. Therefore, because of the comparatively large concentration of dissolved solids and the identification of random values of high concentrations of trace elements, methods for disposal of these materials to prevent their access to public water supplies were assessed.

Results

Operational Modes

By the end of 1977, SO_2 scrubbers were operating at 22 power stations having a scrubbing capacity of approximately 10,375 MWe; 20 have nonregenerable scrubbers, and 2 have regenerable scrubbers. In the nonregenerable category, 7 stations (scrubbing 4460 MWe) use chemical treatment disposal processes, and 13 stations (scrubbing 5680 MWe) dispose of the sludge untreated. The two regenerable systems have a total capacity of 235 MWe. A breakdown of the disposal modes is as follows:

	Treated, Unlined	Treated, Lined Pond	Untreated, Unlined Pond	Gypsum, Lined Pond	Stabilized,[a] Unlined Pond	Untreated, Solar Evaporation
No. of Plants	6	1	7	1	4	1
Total MWe	4293	167	3250	1420	635	375

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[a] Stabilized, e.g., dewatered with fly ash addition; not necessarily the final disposal mode.

Disposal Alternatives

The general categories of disposal and the considerations required for environmental control are shown in Table II. In each case, seepage of rainwater through the sludge and eventual contamination of groundwater pose an environmental concern for all disposal methods. Runoff is a potential source of environmental pollution for landfill sites because these sites are open and do not necessarily return water to the scrubber. Only in the case of ponding is it clear that the disposal site is not directly amenable to land reclamation efforts, although even in some of these cases it may be possible upon retirement to air-dry, cap, and vegetate the site. Consideration of each of these effects are given in the following discussions.

Ponding. In general, the simplest and the least cost (though not necessarily the most environmentally sound) approach to FGD sludge disposal is ponding. This method requires that, if the pond does not contain a base material considered to be impermeable, a liner must be added to prevent seepage. Operationally, sludge ponds exist today which contain either naturally impermeable soils or clay liners transported and placed in the base and on the slopes of the pond. Because of the highly thixotropic nature of these sludges, ponds

Table II. Environmental Effects of Disposal Alternatives

Type of Disposal	Condition of Waste	Primary Drainage	Environmental Effect		
			Seepage	Runoff	Land Reuse
Pond	Untreated[a]	Supernate	Yes	No	No
	or chemically treated[b]	Supernate	Yes	No	Yes
Basin	Untreated[a] or conditioned[c]	Underdrainage	Yes	No	Yes
Landfill	Conditioned[c] or chemically treated[b]	Runoff	Yes	Yes	Yes

[a] Untreated waste refers to FGD sludges as emitted from primary or secondary dewatering equipment.

[b] Chemically treated sludges refer to the waste treated by one of several commercial processes that make these wastes suitable for landfill disposal.

[c] Conditioned waste refers to sludge treated by techniques other than chemical treatment and includes oxidation to gypsum and dewatering by mixing with dry fly ash or other agents that allow the material to be handled in a manner similar to that for soils.

are nonstructural sites and generally are not considered amenable to reclamation, except possibly in areas of low rainfall and high evaporation. Also, if ponded sludges are not dewatered, larger land areas are needed to contain the material.

Ponding with Underdrainage. This approach to ponding is still under evaluation and is not being used operationally at this time. Underdraining and collection of all seepage for return to the scrubber system maintains control of leachate at all times and has been shown at small scale evaluation sites to produce a material structurally capable of supporting personnel and construction equipment. Present evaluations[1] are being made to determine the feasibility of such an alternative regarding (a) site reclamation and (b) relaxation of requirements on the degree of water-tightness of the base material, inasmuch as no appreciable hydraulic head exists. A site of this type collects rainfall via the seepage system and returns it to the scrubber. As a result it is necessary to limit the size of each disposal basin to maintain an acceptable water balance in the scrubber loop. This would be accomplished by dividing the site into sections of approximately 35 to 50 acres at a depth of about 30 feet. Figure 1 shows an underdrained, untreated pond supporting a general purpose farm tractor within one day after a 3-inch rainfall.

Chemical Treatment. The stabilization of FGD sludges by chemical treatment offers the most positive solution to the disposal problem. It converts the sludge to a structural material; decreases its coefficient of permeability to a range of approximately 10^{-5} to 10^{-7} cm/sec, which as a minimum is one order of magnitude better than untreated sludges; reduces the concentration of salt constituents in the leachate by approximately 50%; is amenable to subgrade or above-grade landfilling; and allows the disposal site to be reclaimable. Chemically treated sludges have not been shown to appreciably reduce concentration of trace elements in leachate, and, even though the concentration of major species is reduced, leaching of chemically treated sites should be avoided unless it can be assured that the leachate can be diluted by local groundwater and streams. A general procedure for managing rainfall runoff from a chemically treated site is to collect the runoff in a peripheral ditch which directs the water to a settling pond. Depending on the quality of the water in this pond, it can be decanted to a stream or returned to the scrubber system.

Chemically treated sludges have solids content of approximately 45 to 65 wt% (or possibly higher depending on the dewatering potential and the treatment process used) and attain load bearing strengths in the range of 75 to 300 psi (5.4 to 21.6 tons/ft^2).

Test ponds containing chemically treated sludges are pictured in Figures 2 and 3.

Gypsum. The forced oxidation of sulfite sludges to gypsum or the production of high sulfate sludge from the use of western coal results in a waste material which is readily dewatered by vacuum filtration or by centrifuging to a solids content in the approximate range of 75 to 85 wt%. Leachate from gypsum is similar to that of sulfite sludges and therefore should be prevented from entering water supplies. Because gypsum tends to form a protective

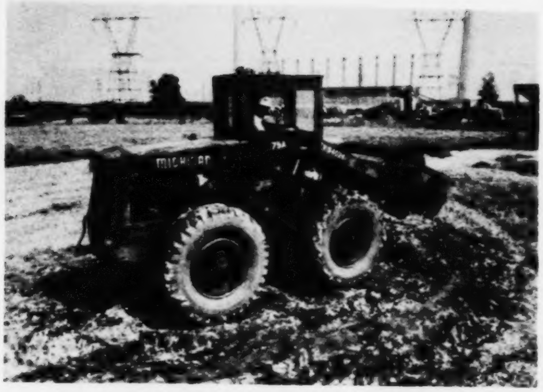

Figure 1. Untreated, unstabilized sludge ponded with underdrainage, one day after 3-inch rainfall.

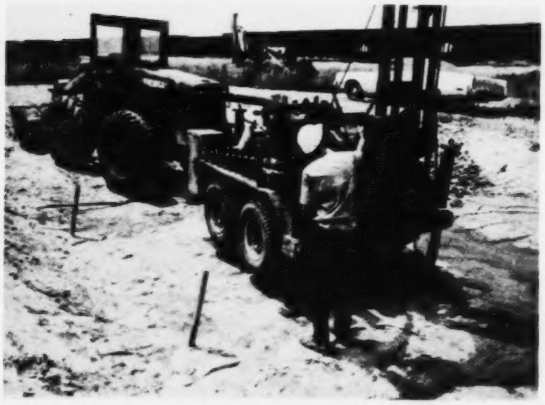

Figure 2. Chemically treated sludge (IU Conversion Systems process).

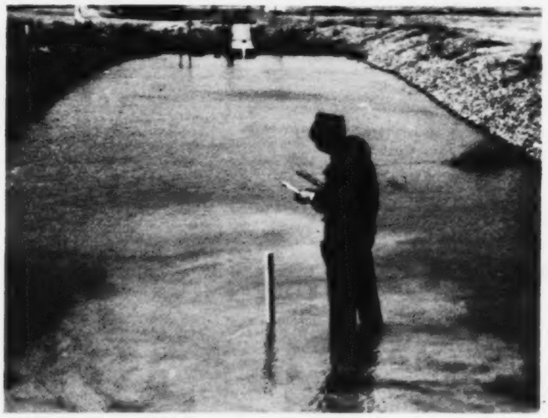

Figure 3. Chemically treated sludge (Dravo process).

surface scale capable of shedding rainwater, tests are currently being conducted to determine the applicability of the disposal of gypsum on the ground without the added benefit of liners or impoundment dikes. Limited results have shown that gypsum sludges crack badly under freeze-thaw conditions, thereby allowing rainwater to enter into the material. Additionally, gypsum sludge slumps in its freshly deposited condition when exposed to rainfall and produces a runoff containing potentially high concentrations of dissolved solids from the sludge, as well as a condition which requires machinery to replace the material on the disposal site. These preliminary results indicate that considerable site maintenance may be required on an operational scale to reconfigure the disposal pile after weathering (freeze-thaw and erosion) and to control the runoff. Tests are continuing for the determination of what control (if any) should be exercised at the site during and after disposal. A gypsum test pile before and after weathering is pictured in Figures 4 and 5.

Sludge Volume Prediction

Landfill volume requirements are strongly affected by the solids content of sludges. A comparative analysis of sludge production in acre feet annually is shown for a 500-MW plant in Figure 6. (This figure neglects the approximate 25% increase in acreage requirements to account for berm slopes and access roads.) If it is assumed that an untreated sludge settles to approximately 50% solids, the acre feet produced in one year for this case would be 250. The advantage for gypsum in this regard (neglecting other environmental factors) would be that approximately 155 acre ft would be produced, providing that the sludge is dewatered to a solids content of 80%. In the case of chemical treatment, if it is assumed that the material is disposed of at a solids content in the range of 60 to 70%, the volume to be disposed of would be in the range of 165 to 190 acre ft.

Disposal Cost Estimates

Cost estimates for ponding and chemical treatment for landfilling have been made and reported by The Aerospace Corporation on several occasions. During studies associated with the EPA Shawnee field disposal evaluation project, Aerospace cost estimates were made of chemical treatment disposal and were reported in the initial report on that study.[2] The Aerospace estimates for lined-pond costs were presented in the initial and second progress reports on sludge disposal[3,4] and at EPA flue gas desulfurization symposiums.[5,6,7] All estimates have been updated in a report to EPA[8] on new source performance standards on a July 1977 basis; these cost estimates are summarized in Table III.

Conclusions

Constituent concentrations of FGD sludges require disposal controls to prevent direct discharge, seepage, or runoff to water supplies. The methods used operationally today are (a) disposal of untreated sludges in ponds with highly impermeable liners or (b) chemical treatment prior to sub-grade or above-grade landfilling. Other methods being evaluated are (a) disposal of untreated sludges in ponds equipped with underdrainage and (b) conversion to gypsum for disposal.

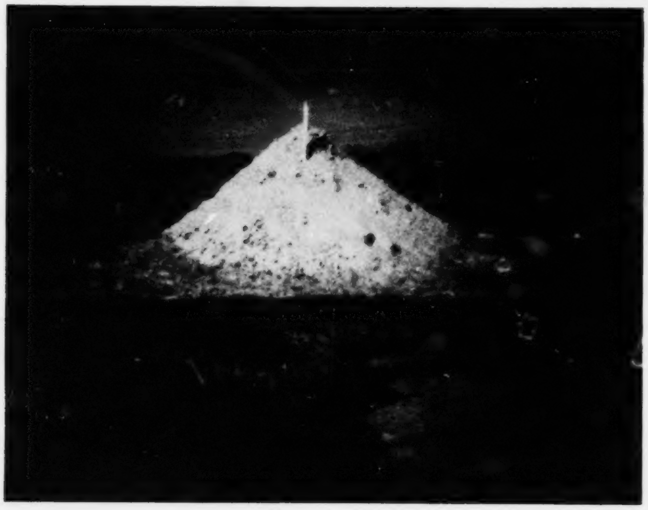

Figure 4. Gypsum filter cake immediately after placement, September 1977, Paducah, Kentucky.

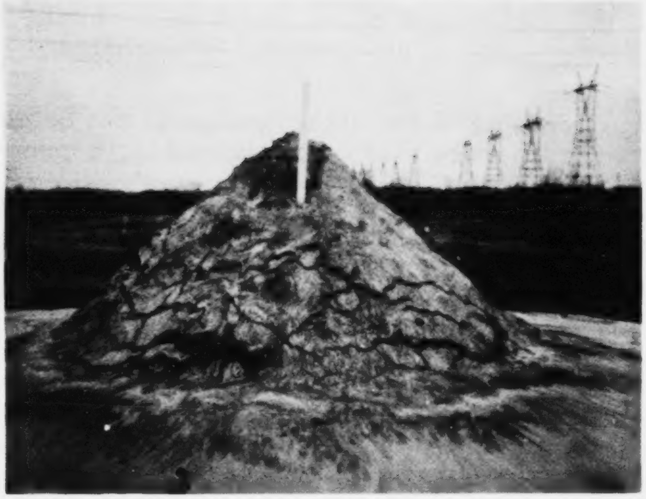

Figure 5. Gypsum filter cake after first winter season, March 1978, Paducah, Kentucky.

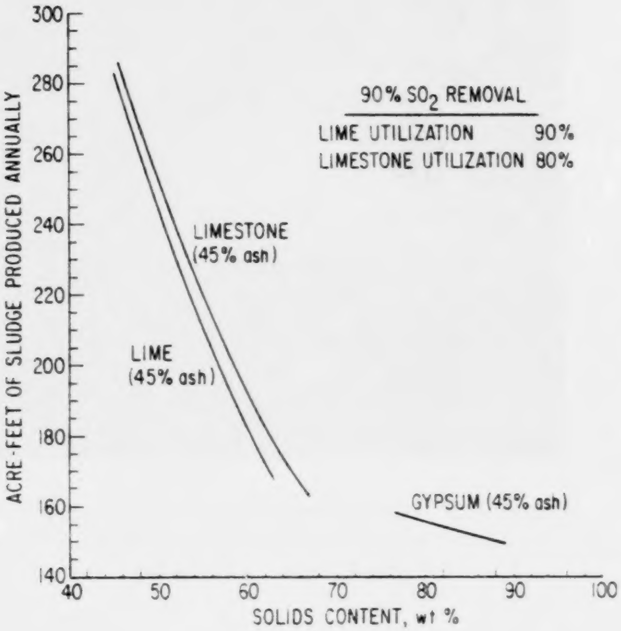

Figure 6. Sludge produced annually (500-MW plant, 3.5% sulfur coal, 12,000 Btu/lb, 14% ash).

Untreated sludge ponds have the disadvantage of not being reclaimable. Those equipped with underdrainage may be reclaimable, depending on evaluations now in progress.

Chemical treatment improves the impermeability of sludges by one order of magnitude or more, reduces the dissolved solids concentration by about 50%, and attains a bearing strength greater than 5 tons/ft^2. Chemically treated sites must be maintained to control seepage or runoff, depending on the process used. Above-grade sites generally require maintenance for runoff control only.

Gypsum sludges dewater readily to 75 to 85 wt% solids. These materials when stacked have exhibited severe surface cracks after freeze-thaw cycling. Therefore, piling or stacking gypsum without considerable site maintenance may not be a feasible disposal method, on the basis of preliminary field tests. Further testing is under way.

Volume production for a 500-MW eastern plant, on the average, is approximately 250, 175, and 155 acre ft annually for untreated, chemically treated, and gypsum sludges, respectively. For a 1000-MW plant, these values would be increased by about 93%. Landfill requirements for these volumes are increased by approximately 25% to account for berm slopes and access roads.

Disposal cost estimates in mills per kilowatt hour (July 1977 dollars) for ponding on indigenous clay, ponding with liner added, and chemical treatment are 0.55, 0.80, and 1.05, respectively, for a 1000-MW plant burning typical eastern coal.

Table III. Disposal Cost Comparison[a]

Cost Basis, Mid-1977 $	Ponding		Landfill, Chemical Treatment	Gypsum
	Indigenous Clay	Liner Added		
Mills/kWh	0.55	0.80	1.05	1.10
$/ton of sludge (dry)	4.90	7.25	9.70	10.30[b]
$/ton of coal	1.50	2.20	2.95	3.10

[a] Notes:

Dollar base: July 1977

Plant characteristics: 1000 MW, 8700 Btu/kWh (0.73-lb coal/kWh)

Coal burned: 3.5% sulfur, 12,000 Btu/lb, 14% ash

Annual average operating hours: 4380 hr/yr (30-yr average)

Plant and disposal site lifetime: 30 yr

SO_2 removal, with limestone absorbent: 90%

Limestone utilization: 80% of all cases except for gypsum, which is 100%

Sludge generated: 4.8×10^5 short tons/yr untreated waste (dry) including ash

Average annual capital charges, 30-yr average: 18% of total capital investment

Cost of land used for disposal: $5000/acre; all land assumed purchased initially; sludge depth, 30 ft

Land depreciation: Total depreciation in 30-yr; straight-line basis

Disposal site: Within one mile of the plant

[b] Cost of forced oxidation and disposal of gysum sludge converted to cost/ton of equivalent quantity of nonoxidized sludge. Divided by 1.08 to convert to gypsum cost. Includes fly ash; disposal is in an indigenous clay pond.

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COMPARATIVE ECONOMICS OF FGD WASTE DISPOSAL

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COMPARATIVE ECONOMICS OF FGD WASTE DISPOSAL

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ABSTRACT

Several series of studies to evaluate the economics of various systems associated with the control of fly ash and sulfur dioxide emissions from power plant flue gases are being conducted by the Tennessee Valley Authority (TVA) for the U.S. Environmental Protection Agency (EPA). One group of studies involves the preparation of economics for the comparison of flue gas desulfurization (FGD) sludge disposal alternatives. Two studies are complete--one report is published and one report is being reviewed by EPA before publication--and a third study is underway. The results of the two completed studies are described in this report.

Six disposal alternatives have been evaluated to date. A base case for each process was established and complete conceptual designs of the systems were prepared for use as a cost estimating basis. Cost estimates and conceptual designs are based on common premises used for all TVA-EPA studies.

The six alternatives evaluated are (1) untreated ponding, (2) Dravo Corporation's process, (3) Chemfix process, (4) IU Conversion Systems' process, (5) untreated sludge - fly ash blending, and (6) gypsum landfill. For each alternative total capital investments and annual revenue requirements were estimated for the base case and major case variations.

COMPARATIVE ECONOMICS OF FGD WASTE DISPOSAL

INTRODUCTION

The U.S. Environmental Protection Agency (EPA) is sponsoring an extensive research and development program to evaluate, develop, and demonstrate sludge disposal alternatives that are environmentally and economically acceptable to the utility industry for flue gas desulfurization (FGD) sludge (1). A major program area that involves the field testing of potential processes for commercial-scale use is The Aerospace Corporation's study being conducted at the Shawnee power plant of the Tennessee Valley Authority (TVA). All of the alternatives evaluated at Shawnee are also considered in the TVA economic studies for sludge disposal options.

Two general categories of FGD processes are available for use by the utility industry: nonregenerable or throwaway processes which produce a waste material for disposal and regenerable or recovery processes that produce a saleable byproduct. Many processes are available in both categories; however, most utilities are selecting the lime or limestone process which produces a throwaway sludge (2). Two categories of waste disposal processes are being used: wet and dry. Wet processes normally involve pond disposal and dry processes usually involve landfill of sludge (3,4). The alternatives evaluated by TVA are representative of a range of disposal options and include both wet and dry disposal processes.

The six alternatives evaluated are (1) untreated ponding, (2) Dravo Corporation's process, (3) Chemfix process, (4) IU Conversion Systems, Inc., (IUCS) process, (5) untreated sludge - fly ash blending, and (6) gypsum disposal. For each process considered, a base case was established and definitive estimates of total capital investments and total annual revenue requirements were calculated. The estimates were all made using a set of carefully defined common premises and are directly comparable. Cost estimates are based on process background information, flowsheets, material and energy balances, equipment and system requirements, and raw material, labor, and utility costs. All estimates of capital investment are projected to mid-1979 and revenue requirements to mid-1980.

BACKGROUND AND DESCRIPTION OF ALTERNATIVES

Many power plants with nonregenerable FGD systems that are now in operation in the United States use a sludge disposal method involving some form of onsite ponding or impoundment of untreated material. This method of disposal, although popular, will not necessarily be the best option for future installations. Drawbacks, such as potential disposal regulations and limited land availability, have made necessary the development of other disposal options for FGD wastes (1). Several other treatment options involving chemical and physical stabilization are available to the utility industry.

The technology associated with the six disposal alternatives evaluated in the TVA-EPA work and discussed in this paper is assumed to be proven, but in many cases is in the development stage and is not actually proven in full-scale application (i.e., forced oxidation, simultaneous sulfur dioxide (SO_2) and fly ash removal, etc.). The primary emphasis of the work was to evaluate the economics of the disposal alternatives rather than the process technology (5).

Untreated Ponding

As stated earlier, the untreated ponding option is the alternative selected most often by the utility industry. Effluent from the scrubber system is pumped directly to a pond and allowed to settle. Excess water is recycled to the scrubber system. Very few items of equipment are required if this option is used, but the capital investment for the disposal pond is very high (7).

Dravo Process

Dravo offers two basic processes for FGD sludge disposal (pond or landfill). Although the pond or impoundment alternative was the base case for TVA studies, the more recently promoted landfill process may be more economically attractive. Effluent from the scrubber system is partially dewatered using a thickener before mixing with Dravo's fixation additives (Thiosorbic lime and Calcilox). The treated material is then pumped to an impoundment area where the material settled and is eventually stabilized. Excess water is recycled to the scrubber system. Dravo's fixation agents and their entire fixation process are patented (Synearth process) (6).

IUCS Process

The IUCS system is called the Poz-O-Tec process and involves chemical stabilization of calcium-based waste materials by mixing with lime and fly ash. Scrubber system effluent is dewatered using a thickener and rotary drum filter. The dewatered material (containing about 60% solids) is then mixed with fixation additives (lime and fly ash) and trucked to a disposal site for landfill disposal. Fly ash is a necessary ingredient for stabilization and can be blended with the sludge cake at the additive mixing stage of processing or included with the sludge following removal simultaneously with the SO_2 in the precooling and scrubbing stages of the FGD process. IUCS reports that the stabilized material is claylike and can be easily handled, transported by truck, placed, and compacted, and that the landfill is structurally suitable for future reclamation (7).

Chemfix Process

Chemfix has been applying their technology for sludge stabilization to wastes generated by metal finishing, automotive assembly, and electronics operations for several years (8). Chemfix offers a process that yields a treated stabilized sludge that is reported to be suitable for landfill disposal. Effluent from the scrubber system is dewatered using a thickener and rotary drum filter. The filter cake is mixed with two chemical additives (Portland cement and sodium silicate), during which stabilization of the sludge is achieved.

Untreated Sludge - Fly Ash Blending

Many power plants are meeting particulate emission requirements for fly ash by installing equipment for dry fly ash collection. Dry fly ash can be used in many cases as an additive for blending with dewatered scrubber sludge to yield a physically stable material. This process of sludge treatment would allow the utility to dispose of both fly ash and scrubber wastes in one operation and also to produce a waste product that is suitable for landfill disposal.

Effluent from the scrubber system is dewatered using a thickener and rotary drum filters. The filter cake is mixed with dry fly ash which is pneumatically conveyed from the fly ash collection system to the sludge disposal facility. The blended material is transported by truck to a landfill disposal site. TVA studies indicate that this procedure can be used to produce a product suitable for landfill disposal and that handling with trucks and earthmoving equipment is feasible (9).

Gypsum

The lime and limestone FGD processes can be modified to include a processing step to force the oxidation of calcium sulfite sludge (the normal product of these processes) to gypsum. Gypsum is a more desirable waste product because of improved settling properties (settling rate is about 10 times greater than $CaSO_3$) and therefore a reduced volume of material can be attained through dewatering. The landfill disposal of gypsum can be accomplished without the use of blending or mixing equipment and fixation additives. Underflow from the scrubber system is dewatered, using a thickener and rotary drum filter, before it is hauled by truck to a landfill disposal site. Tests conducted at Shawnee power plant indicate that SO_2 and fly ash can be removed simultaneously in the scrubber system and therefore the equipment for dry fly ash collection is not needed (10,11).

EVALUATION OF ALTERNATIVES

A comparative economic evaluation of several processes requires that the basis for the capital investment and revenue requirement estimates be the same. All TVA studies are made using a predetermined set of design and economic premises for the power plant, fuel, FGD system, and estimate calculation procedures. These premises that allow the comparison of estimates are summarized in the following paragraphs.

Base Case Design Premises

Power Plant--

1. The plant is newly constructed and has a 30-year life.
2. The single coal-fired unit has an output of 500 MW.
3. The total operating life is 127,500 hours with an average annual capacity of 4,250 hours.
4. The power unit heat input requirement is 9,000 Btu/kWh.
5. The coal heating value is 10,500 Btu/lb.
6. The coal contains 3.5% (by wt) sulfur (dry) and 16% (by wt) ash.

FGD System--

1. A limestone scrubbing process is used for SO_2 removal.
2. SO_2 and fly ash are removed to meet NSPS. [EPA issued Federal Standards of Performance for New Stationary Sources (often called "new source performance standards" or NSPS).] The allowable SO_2 emission is 1.2 lb/MBtu heat input and the particulate emission, 0.1 lb/MBtu heat input.
3. Eighty-five percent of the ash present in the coal is emitted as fly ash.
4. Ninety-five percent of the sulfur in the coal is emitted as SO_2.
5. Effluent from the scrubber system contains 15% solids.
6. All storage facilities have a 30-day capacity and feed bins, intermediate storage tanks, etc., have an 8-hour capacity.

Untreated Ponding--

1. Effluent (15% solids) from the scrubber system is pumped to a clay-lined disposal pond.

2. The pond is located 1 mile from the scrubber facilities.

3. The sludge settles to 50% solids in the pond and excess water is recycled to the scrubber system.

4. The FGD process stoichiometry is 1.5 mols calcium oxide per mol SO_2 removed.

5. Fly ash and SO_2 are removed simultaneously in the scrubber system; therefore the sludge contains both fly ash and calcium wastes.

6. Fifteen percent of the SO_2 removed is converted to gypsum and the remaining 85% calcium sulfite.

Dravo Process--

1. Thickened sludge (35% solids) is treated with Dravo additives: Calcilox (7% of dry solids) and Thiosorbic lime (1% of dry solids).

2. Treated sludge is pumped 1 mile to a clay-lined pond for disposal.

3. Stabilization as a soillike material occurs over a 2- to 4-week period. Fixed sludge is 50% solids and excess water is recycled to the scrubber system.

4. The FGD process stoichiometry is 1.5 mols calcium oxide per mol SO_2 removed.

5. Fly ash and SO_2 are removed simultaneously in the scrubber system; therefore the sludge contains both fly ash and calcium wastes.

6. Fifteen percent of the SO_2 removed is converted to gypsum and the remaining 85% calcium sulfite.

IUCS Process--

1. Dewatered sludge (60% solids) is treated with lime (4% of dry solids).

2. Trucks are used to transport the treated material to a landfill disposal site located 1 mile from the scrubber facilities.

3. Treated sludge is assumed to have claylike properties and can be placed and compacted in a landfill with typical earthmoving equipment.

4. The FGD process stoichiometry is 1.5 mols calcium oxide per mol SO_2 removed.

5. Fly ash and SO_2 are removed simultaneously in the scrubber system; therefore the sludge contains both flyash and calcium wastes.

6. Fifteen percent of SO_2 removed is converted to gypsum and the remaining 85% calcium sulfite.

Chemfix Process--

1. Thickened sludge (35% solids) is transported by pipeline (1 mile) to the disposal site where additional dewatering, mixing with fixation additives, and landfill placement occurs.

2. Dewatered sludge (60% solids) is stabilized by mixing with two Chemfix additives: Portland cement (7% of dry solids) and sodium silicate (2% of dry solids).

3. Treated material is placed and compacted as landfill using typical earthmoving equipment.

4. The FGD process stoichiometry is 1.5 mols calcium oxide per mol SO_2 removed.

5. Fly ash and SO_2 are removed simultaneously in the scrubber system; therefore the sludge contains both fly ash and calcium wastes.

6. Fifteen percent of the SO_2 removed is converted to gypsum and the remaining 85% calcium sulfite.

Untreated Sludge - Fly Ash Blending--

1. Dewatered sludge (60% solids) is blended with dry fly ash to yield a physically stable material.

2. Fly ash is removed from the flue gas to meet NSPS using an electrostatic precipitator (ESP) and pneumatically conveyed to the sludge treatment area for blending with sludge.

3. The blended material (about 75% solids) is transported to a landfill disposal site by truck (1 mile).

4. Typical earthmoving equipment is used for placement and compaction in a landfill.

5. The FGD process stoichiometry is 1.5 mols calcium oxide per mol SO_2 removed.

6. Fifteen percent of the SO_2 removed is converted to gypsum and the remaining 85% calcium sulfite.

Gypsum--

1. The limestone FGD process is modified to provide forced oxidation of calcium sulfite sludge to gypsum. The FGD process stoichiometry is 1.1 mols calcium oxide per mol SO_2 removed.

2. Ninety-five percent of the SO_2 removed is converted to gypsum and the remaining 5% calcium sulfite.

3. Fly ash and SO_2 are removed simultaneously in the scrubber loop to meet NSPS; therefore, sludge contains both gypsum and fly ash.

4. Dewatered gypsum (about 80% solids) is transported by truck (1 mile) to the landfill disposal site.

5. Typical earthmoving equipment is used for placement and compaction of the gypsum in the landfill.

Economic Premises

A midwestern plant location was selected because of coal availability for the large number of coal-fired plants in this region. Other economic assumptions are summarized as follows:

1. All capital cost estimates are based on Chemical Engineering cost indices (labor index - 237.9, material index - 264.9). Capital costs are projected to mid-1979 using these indices. Construction on the project is assumed to have started in mid-1977 and to be completed in mid-1980.

2. Direct capital costs cover process equipment, piping and insulation, transport lines, foundations and structural, excavation and site preparation, roads and railroads, electrical instrumentation, buildings, and trucks and earthmoving equipment. Material and labor (fabrication and installation) costs for each of these items were estimated. These estimates are based on costs obtained from vendors and on related literature information.

3. Indirect capital costs include engineering design and supervision, architect and engineering contractor expenses, construction expenses, contractor fees, contingency, allowance for startup and modifications, and interest during construction. Two other capital costs not included as indirect costs, but in the total capital investment, are working capital and land. These estimates are based on current industry practice and authoritative literature sources.

4. Direct costs for revenue requirements include raw materials, labor, electricity, equipment fuel and maintenance, and analyses. These costs are projected to mid-1980.

5. Indirect costs for revenue requirements are capital charges and overheads.

6. Capital charges are based on regulated utility economics.

7. Revenue requirements are projected for an annual 7000 hr/yr (first year) operation. Other estimates are made for lifetime revenue requirements that are based on the declining operating profile of the plant.

Case Variations in Design Premises

The base case design premises were altered for selected variables in order to evaluate the effects of changes in operating conditions and site-specific design factors. Several of the variations which were considered are as follows:

1. Plant size: 200 and 1500 MW (the 1500-MW plant is assumed to be three 500-MW units).
2. Coal composition: Sulfur content, 2.0% and 5.0%; ash content, 12% and 20%.
3. Remaining life of an existing plant: 25, 20, and 15 years.
4. Distance to disposal: 5 and 10 miles.
5. Availability of land for disposal site. Construction: 50% and 75% of optimum.

RESULTS

Two TVA-EPA studies to evaluate the economics of six FGD sludge disposal alternatives are complete. The capital investments and revenue requirements of the base cases and major case variations for the six options are discussed in this paper. Additional details concerning the cost estimates can be obtained by reviewing the two TVA-EPA reports (1) (one of the two reports is not yet published, but details are available from the author of this paper).

Total System Costs

The total cost of SO_2 and particulate emission control can be obtained by combining the cost estimates of the FGD system with waste disposal system costs. Estimates of FGD system costs (total capital investment and annual revenue requirements) are available from other TVA-EPA studies (5,13) and are suitable for combining with the waste disposal system costs discussed in this report. These costs are summarized in Tables 1 and 2. The FGD costs presented in this paper apply only to the base case conditions and therefore cannot be used with waste disposal systems other than the base cases.

Unit Revenue Requirements

Unit revenue requirements for the base case system and several major case variations are shown in Table 3.

TABLE 1. SUMMARY OF CAPITAL INVESTMENTS FOR COMBINED FGD AND SLUDGE DISPOSAL SYSTEMS

Disposal process[a]		Total capital investments[b]					
		FGD system		Disposal system		Combined system	
Variation	Site description	k$	$/kW	k$	$/kW	k$	$/kW
Untreated	Ponding	36,368[c]	72.8	17,211	34.4	53,579	107.2
Dravo	Ponding	36,368[c]	72.8	24,114	48.2	60,482	121.0
Dravo	Landfill	36,368[c]	72.8	12,670	25.3	49,038	98.1
Chemfix	Landfill	36,368[c]	72.8	13,531	27.2	49,899	99.8
IUCS	Landfill	36,368[c]	72.8	10,717	21.4	47,085	94.2
Gypsum	Landfill	38,671[c,d]	77.3	5,411	10.7	44,082	88.2
Untreated sludge - fly ash blending	Landfill	45,982[e]	92.0	8,605	17.2	54,587	109.2

a. Dewatering equipment for all cases included in the disposal system.
b. The amounts shown are for the base case (mid-1979 costs).
c. Costs are for an FGD system which removes both SO_2 and fly ash in the scrubber loop.
d. Cost includes additional equipment required for forced oxidation ($2,300,000).
e. An electrostatic precipitator (ESP) is used to remove fly ash and its installed cost is included ($9,614,000).

TABLE 2. SUMMARY OF REVENUE REQUIREMENTS FOR

COMBINED FGD AND SLUDGE DISPOSAL SYSTEMS

Disposal process		Revenue requirements[a]						
		FGD system		Disposal system			Combined systems	
Variation	Site description	Total annual $	Mills/ kWh	Total annual $	Mills/ kWh	$/ton dry solids	Total annual $	Mills/ kWh
Untreated	Ponding	11,841,500[b]	3.38	3,280,000	0.94	8.08	15,121,500	4.32
Dravo	Ponding	11,841,500[b]	3.38	6,701,000	1.91	15.32	18,542,500	5.30
Dravo	Landfill	11,841,500[b]	3.38	6,620,000	1.89	15.16	18,461,500	5.27
Chemfix	Landfill	11,841,500[b]	3.38	6,988,000	2.00	16.51	18,829,500	5.38
IUCS	Landfill	11,841,500[b]	3.38	5,291,000	1.51	12.55	17,132,500	4.90
Gypsum	Landfill	12,846,800[b,c]	3.67	3,117,500	0.89	7.86	15,964,300	4.56
Untreated sludge - fly ash blending	Landfill	13,816,500[d]	3.94	3,735,000	1.07	9.20	17,551,500	5.01

[a] The amounts shown are for the base case (mid-1980 costs).

[b] Costs are for an FGD system which removes both SO_2 and fly ash in the scrubber loop.

[c] Cost includes that associated with forced oxidation equipment ($1,005,300).

[d] An ESP is used to remove fly ash and its associated operating costs are included ($1,975,000).

TABLE 3. UNIT REVENUE REQUIREMENTS - ALL PROCESSES[a]

	Disposal process											
	Untreated		Dravo		IUCS		Chemfix		Untreated sludge - fly ash blending		Gypsum	
	Mills/ kWh	$/dry ton	Mills/ kWh	$/dry ton	Mills/ kWh	$/dry ton	Mills/ kWh	$/dry ton	Mills/ kWh	$/dry ton	Mills/ kWh	$/dry ton
Base case[b]	0.94	8.08	1.91	15.32	1.51	12.55	2.00	16.51	1.07	9.20	0.89	7.86
Variation from base case												
200 MW	1.44	12.12	2.60	20.41	2.55	20.68	3.24	26.14	1.96	16.51	1.79	15.42
1500 MW	0.64	5.55	1.36	10.87	0.99	8.23	1.37	11.31	0.65	5.64	0.47	4.17
Existing, 25-year life	0.55	4.69	1.32	10.30	1.01	8.24	1.40	11.36	1.07	9.01	0.88	7.63
Existing, 20-year life	0.45	3.80	1.21	9.50	1.02	8.26	1.41	11.39	1.06	8.97	0.88	7.62
Existing, 15-year life	0.38	3.19	1.16	9.04	1.04	8.43	1.43	11.59	1.06	8.94	0.88	7.61
12% ash in coal	0.83	8.68	1.69	16.43	1.30	13.05	1.78	17.86	1.02	10.77	0.86	9.23
20% ash in coal	1.03	7.44	2.12	15.58	1.71	11.84	2.17	15.00	1.11	8.03	0.92	6.75
2% sulfur in coal	0.75	9.37	1.52	17.45	1.33	15.57	1.70	20.19	0.91	11.26	0.77	9.74
5% sulfur in coal	1.10	7.35	2.29	14.08	1.77	11.29	2.36	15.01	1.20	7.88	0.93	6.45
5 miles to disposal	1.58	13.61	2.32	18.57	1.85	15.40	2.48	20.49	1.25	10.81	1.06	9.37
10 miles to disposal	2.14	18.48	2.67	21.39	2.14	17.73	2.86	23.63	1.39	11.96	1.22	10.80
Constrained acreage (50% of optimum)	1.18	10.15	2.60	20.82								
Constrained acreage (75% of optimum)	0.96	8.29	2.25	18.06								

a. Basis

Midwest plant location, mid-1980 costs; 7,000 hr/yr plant on-stream time; SO_2 and fly ash removed to meet NSPS.

b. Base case

New 500-MW plant with 30-year life.

Sludge Disposal System Costs

Major case variations and their effects on costs are discussed in the following section of the paper. The costs shown in these tables represent only the costs associated with the sludge disposal area.

Power Plant Size--

The power plant size has an almost direct effect on sludge disposal costs. A slight economy of scale is seen for the plant sizes evaluated. Table 4 is a summary of sludge disposal process costs for the alternatives evaluated.

Coal Composition--

The sulfur and ash contents of the coal also have a direct effect on the quantity of sludge for disposal. Cost estimates were made for variable sulfur and ash percentages of the coal. These estimates are summarized in Table 5.

Remaining Plant Life--

In many cases existing power plants (5-15 years old) are required to install FGD systems to meet emission regulations. Several cost estimates were made to evaluate the sludge disposal costs for plants with remaining operating times of less than 30 years (15, 20, and 25 years). Capital investments for these cases were considerably less if the disposal alternative involved ponding. Unit revenue requirements were increased because the depreciation of capital was taken over a shorter period of time. Table 6 summarizes the remaining life case variation estimates.

Distance to Disposal Site--

Case variations were considered to determine the effect of the distance to the waste disposal site on capital investment and revenue requirements. These results are summarized in Table 7. The capital investment and revenue requirements increase rapidly with the increasing distance to disposal for alternatives using pipelines for slurry transport. Costs increase for alternatives for using trucks for transport, but not as much as the pipeline transport alternatives.

Availability of Land--

The quantity of land available for construction of a disposal pond for untreated sludge can be a significant factor in selecting a disposal alternative. Several cost estimates were made to evaluate the effect of land availability on costs. Estimates are normally made in TVA studies by determining the minimum total pond cost by optimizing between land cost and construction costs. The quantity of land is therefore the amount that should be used to obtain the lowest overall pond cost.

TABLE 4. TOTAL CAPITAL INVESTMENTS AND ANNUAL REVENUE REQUIREMENTS

FOR PLANT SIZE CASE VARIATIONS

	Total capital investment, k\$[a]			Annual revenue requirements, k\$		
	Power plant size, MW			Power plant size, MW		
Disposal process	200[b]	500[c]	1500[b]	200[b]	500[c]	1500[b]
Untreated	9,800	17,211	36,455	2,014	3,280	6,746
Dravo	13,942	24,114	48,235	3,643	6,701	14,264
IUCS	7,193	10,717	20,105	3,567	5,291	10,411
Chemfix	9,259	13,531	24,104	4,529	6,988	14,362
Untreated sludge - fly ash blending	6,126	8,605	18,282	2,742	3,735	6,867
Gypsum	3,988	5,411	9,826	2,502	3,118	4,961

a. New plant with 30-year life; Midwest plant location; mid-1979 capital costs; mid-1980 revenue requirements; 7,000 hr/yr on-stream time; coal analyses (by wt): 3.5% sulfur (dry basis), 16% ash; fly ash and SO_2 removed to meet NSPS; 1 mile to disposal site.

b. Base case premises except plant size.

c. Base case.

TABLE 5. TOTAL CAPITAL INVESTMENT AND ANNUAL REVENUE REQUIREMENTS FOR COAL COMPOSITION CASE VARIATIONS

Disposal process	Total capital investment, k$[a]				Annual revenue requirement, k$[a]			
	Sulfur in coal, %		Ash in coal, %		Sulfur in coal, %		Ash in coal, %	
	2[b]	5[c]	12[d]	20[e]	2[b]	5[c]	12[d]	20[e]
Untreated	13,390	20,655	15,031	19,055	2,639	3,869	2,902	3,609
Dravo	19,251	28,523	21,466	26,028	5,314	8,007	5,924	7,406
IUCS	9,345	11,957	9,025	12,283	4,654	6,118	4,533	5,971
Chemfix	11,879	14,192	11,123	14,854	5,935	8,263	6,229	7,600
Untreated sludge - fly ash blending	7,356	9,534	7,917	9,309	3,186	4,199	3,581	3,896
Gypsum	4,782	5,884	5,042	5,707	2,707	3,252	3,018	3,206

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a. New plant with 30-year life; Midwest plant location; mid-1980 operating costs; mid-1979 capital costs; 7,000 hr/yr on-stream time; fly ash and SO_2 removed to meet NSPS; 1 mile to disposal site.

b. Base case premises except percent sulfur in coal and coal heating value (10,700 Btu/lb).

c. Base case premises except percent sulfur in coal and coal heating value (10,400 Btu/lb).

d. Base case premises except percent ash in coal and coal heating value (11,100 Btu/lb).

e. Base case premises except percent ash in coal and coal heating value (9,900 Btu/lb).

TABLE 6. TOTAL CAPITAL INVESTMENT AND ANNUAL REVENUE REQUIREMENT

FOR REMAINING POWER PLANT LIFE CASE VARIATIONS

	Total capital investment, k$[a]				Annual revenue requirements, k$[a]			
	Remaining power plant life, year				Remaining power plant life, year			
Disposal process	30[b]	25[c]	20[c]	15[c]	30[b]	25[c]	20[c]	15[c]
Untreated	17,211	14,578	11,399	8,822	3,280	2,906	2,135	2,130
Dravo	24,114	21,416	18,281	15,553	6,701	6,377	5,941	5,728
IUCS	10,717	10,591	10,402	10,269	5,291	5,402	5,430	5,559
Chemfix	13,531	13,400	13,204	13,077	6,988	7,152	7,191	7,359
Untreated sludge - fly ash blending	8,605	8,528	8,381	8,276	3,735	3,739	3,724	3,712
Gypsum	5,411	5,174	5,115	5,076	3,118	3,097	3,091	3,087

a. Midwest plant location; mid-1979 capital costs; mid-1980 revenue requirements; 7,000 hr/yr on-stream time; fly ash and SO_2 removed to meet NSPS; 1 mile to disposal site.

b. Base case.

c. Same as base case except remaining plant life and boiler heat rate (9,200 Btu/kWh).

TABLE 7. TOTAL CAPITAL INVESTMENT AND ANNUAL REVENUE REQUIREMENTS FOR DISTANCE TO DISPOSAL SITE CASE VARIATIONS

Disposal process	Total capital investment, k$[a]			Annual revenue requirement, k$[a]		
	Distance to disposal site, mile			Distance to disposal site, mile		
	1[b]	5[c]	10[c]	1[b]	5[c]	10[c]
Untreated	17,211	26,836	37,420	3,280	5,527	7,504
Dravo	24,114	30,994	37,765	6,701	8,124	9,360
IUCS	10,717	11,377	11,891	5,291	6,490	7,475
Chemfix	13,531	18,313	20,227	6,988	8,675	10,003
Untreated sludge - fly ash blending	8,605	8,969	9,334	3,735	4,389	4,855
Gypsum	5,411	5,750	6,514	3,118	3,719	4,286

a. New plant with 30-year life; Midwest plant location; mid-1979 capital costs; mid-1980 revenue requirements; 7,000 hr/yr on-stream time; fly ash and SO_2 removed to meet NSPS.

b. Base case.

c. Same as base case except distance to disposal site.

Estimates shown in Table 8 are for systems with disposal ponds constructed on a less than optimum acreage. Although total land costs are less for these cases, pond construction costs are much higher than for the optimum ponds.

Other Variations--

Several other variations from the base case design and economic premises were considered in TVA sludge studies. Since these case variations had a lesser effect on the costs than the variations discussed, the results are not included in this paper.

Lifetime Revenue Requirements

Estimates of the total revenue requirements of waste disposal processes over the 30-year system life were estimated. These costs, as shown in Table 9, are cumulative over the 30-year plant life.

CONCLUSIONS

Several conclusions can be derived from the results generated by the TVA-EPA sludge disposal economic studies.

1. The base case sludge disposal system requiring the lowest capital investment and annual revenue requirement was gypsum disposal. This alternative requires a much smaller investment for equipment than any other alternative except untreated ponding which requires a very expensive disposal pond. The selection of this alternative would require that a typical limestone FGD system be modified to include the forced oxidation of sulfite (SO_3) compounds to gypsum. This requires an additional capital investment of $2,300,000.

2. In all case variations, the gypsum process had the lowest total capital investment and annual revenue requirements.

3. The alternatives involving pond disposal (untreated and Dravo) required the highest capital investments. All other processes were for landfill disposal.

4. The three processes involving chemical treatment (Dravo, IUCS, and Chemfix) all had higher annual revenue requirements than the three processes involving no chemical treatment.

5. Both unit capital investment and unit revenue requirements were slightly lower for large plant size.

6. Capital requirements and revenue requirements vary almost directly in proportion to the quantity of sludge for disposal. A slight economy of scale is seen. Cases involving coal, ash, and sulfur content variations are examples of this effect.

TABLE 8. TOTAL CAPITAL INVESTMENTS AND

ANNUAL REVENUE REQUIREMENTS FOR LAND AVAILABILITY

CASE VARIATIONS FOR UNTREATED PONDING DISPOSAL

Case variation	Land requirement, acre	Total capital investments, k$[a]	Annual revenue requirements, k$[b]
Optimum land[c]	407	17,211	3,280
75% optimum land[d]	305	17,985	3,365
50% optimum land[d]	204	22,676	4,119

a. New 500-MW plant with 30-year life; Midwest plant location; mid-1979 costs; fly ash and SO_2 removed to meet NSPS; 1 mile to disposal site.

b. Same as footnote "a" except costs are mid-1980.

c. Base case for untreated disposal option.

d. Same as base case except acreage and cost of disposal pond.

TABLE 9. SUMMARY OF LIFETIME REVENUE REQUIREMENTS FOR ALL PROCESSES[a]

Disposal process	Actual cumulative lifetime revenue requirements, $	Lifetime average unit revenue requirements, mills/kWh	Discounted cumulative lifetime revenue requirements, $[b]	Levelized unit revenue requirements, mills/kWh[c]
Untreated	97,757,800	1.53	33,612,100	1.35
Dravo	175,764,900	2.76	62,052,600	2.50
IUCS	131,224,200	2.06	45,381,700	1.83
Chemfix	167,942,300	2.63	59,099,300	2.38
Untreated sludge - fly ash blending	96,526,800	1.51	32,801,900	1.32
Gypsum	78,072,400	1.22	216,513,400	1.07

a. Basis

New plant with 30-year life; Midwest plant location; mid-1980 costs; fly ash and SO_2 removed to meet NSPS; operating profile: 7,000 hr/yr for 10 years, 5,000 hr/yr for 5 years, 3,500 hr/yr for 5 years, 1,500 hr/yr for 10 years; coal analysis (wt %) - 3.5% sulfur (dry), 16% ash.

b. Discounted to initial year at 10%.

c. Equivalent to discounted process cost over life of power plant.

7. The remaining life of a power plant has a significant effect on the relative ranking of capital investments for the two alternatives involving pond disposal. As the plant life is reduced, these alternatives become more favorable.

8. The distance to the disposal site greatly increases the capital investments for the untreated, Dravo, and Chemfix alternatives. These increases are primarily due to the additional costs for pumps and pipelines (other alternatives involve truck transportation to disposal site).

9. Case variations for disposal of untreated sludge in ponds constructed on less than the optimum acreage have higher total capital investments than the base (optimum acreage) case. These variations illustrate the potential problems for plants with a limited quantity of land available for pond construction.

10. Alternatives involving truck transport and landfill disposal generally had higher revenue requirements, but lower capital investments than the alternatives involving pipeline transport and pond disposal.

The results presented in this paper do not take into account site-specific waste disposal conditions that a utility may encounter when selecting a system for installation. Results are based only on predetermined design and economic premises and should not be interpreted to represent a site-specific disposal situation.

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